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British Columbia

Ministry of
Environment

POLLUTION CONTROL ASPECTS OF NATURAL GAS PROCESSING AND THE INDUSTRY IN BRITISH COLUMBIA

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1.0 INTRODUCTION

Natural gas used by consumers is generally considered as a "clean" fuel. It is rare to find such "clean" fuel directly from a gas reservoir. As the reservoir gas contains, in most cases, certain undesirable constituents, some processing of the gas is necessary to meet specifications for sales gas quality. The extent of processing depends on the nature as well as the quantity of the impurities present in the gas. Most common impurities present in the reservoir gas are water, liquid hydrocarbons, carbon dioxide and sulphurous gases mainly in the form of hydrogen sulphide. In the industry carbon dioxide and hydrogen sulphide are called acid gases, and a reservoir gas containing these species is referred to as sour gas. The sales gas, commonly called sweet gas, contains only a trace amount of sulphurous odourant, equivalent to approximately 4 ppm_v of hydrogen sulphide.

The gas processing technology is well established except for the fact that because of environmental considerations improved and new processes are being developed to comply with pollution regulations. A number of waste streams are generated during the processing of the raw gas which also require certain treatment prior to discharge to the receiving environment.

In this report an attempt has been made to provide information on the pollution control aspects of natural gas processing after a brief history of the industry and an outline of a typical plant.

2. HISTORY(1,2)

The history of natural gas processing industry in Canada began in 1924 in southwestern Ontario when Union

Gas Company started desulphurizing the natural gas from Tilbury field by Koppers process at Port Alma(1). The gas in the Tilbury area was discovered in 1905, and although it contained some H_2S, the gas was delivered unprocessed to the nearby town of Chatham by the end of 1906 after precautions were taken to ensure a minimum pipe leakage and adequate ventilation for burners. In 1913 when the company started delivering the gas to Brantford and other nearby towns, there were strong protests about the obnoxious odour. That led the company to decide on desulphurization of the entire gas supply at a central plant. Until then some isolated purifiers had been installed for specific requirements. Reportedly the company attempted to operate a gas purification plant at Glenwood but had abandoned that in favour of a large Koppers plant at Port Alma. The Port Alma facility operated the Koppers plant until 1950 when it was replaced by a monoethanolamine absorption plant.

In western Canada the natural gas industry started with the production at Royalite No. 4 well of the Turner Valley field in Alberta in late 1924. Initially the emphasis was given on separation of the high proportion of natural gasoline from the well gas at wellhead, as the demand for gas was not much at that time. Subsequently central gas processing plants were built at Turner Valley village in 1933 and near Hartell and Longview by 1936. Until 1947 these three plants comprised the natural gas processing industry in western Canada, and since then numerous plants have been built as the exploration and production of gas fields mushroomed.

In British Columbia the industry began in 1957 when Westcoast Transmission Company started its first gas processing plant at Taylor. Since then two more major gas processing plants have been built in the northeastern part of the province to process additional gas supply from newly discovered fields. These plants are listed in Table 1, and their locations are shown in Figure 1 along with gathering fields and transmission pipelines.

3. NATURAL GAS PESERVES IN BRITISH COLUMBIA(2)

In the province of British Columbia there are nine major sedimentary basins (Figure 2) where potential exists for natural gas fields. The exploration to date has demonstrated the presence of natural gas in the Western Canadian and Fernie basins, although some drilling has been carried out in other basins except the Whitehorse. While there have been significant discoveries and production of sales gas in the Western Canadian basin (Figure 3), only small volumes of natural gas have been found in the Fernie basin. The estimates of natural gas resources in the province according to the Govier Commission Report are shown in Table 2.

4. CHARACTERISTICS OF NATURAL GAS

As with any other naturally occurring substances, it is extremely rare to find a source of pure natural gas ready for use by consumers. The gas in a natural reservoir is normally held mixed with some liquid hydrocarbons and water. Under reservoir conditions, at pressures of 6 895 kPa or more, these are mostly present as vapours. However, with a few exceptions some of these substances condense out in the well or at the well-head during gas extraction. In the industry the term condensate is commonly used for the hydrocarbon fractions of the substances referred to above. The condensate may contain some crude oil, lighter hydrocarbons such as propane and butane, and others grouped as "pentane+".

TABLE 1

EXISTING NATURAL GAS PROCESSING PLANTS IN BRITISH COLUMBIA

		Daily Production Capacity				
Company	Location	Raw Gas $m^3 \times 10^6$	Sales Gas $m^3 \times 10^6$	Natural gas Liquids* m^3	Sulphur t	Remarks
Major Plants						
Petro-Canada Exploration Inc. and Westcoast Transmission Co. Ltd.	Taylor	15.5	15.1	795	467	
Westcoast Transmission Co. Ltd.	Fort Nelson	28.3	22.6		670	
Westcoast Transmission Co. Ltd.	Pine River (Chetwynd)	7.4	5.8		1060	
Other Plants.						
Amoco Canada Petroleum Co. Ltd.	Beaver River	1.72	-		-	Sent to Fort Nelson plant
B.P. Exploration Canada Ltd.	Bullmoose	2.01	-		-	Sent to Pine River Plant.
B.P. Exploration Canada Ltd.	Sukunka	2.58	-		24	Sent to Pine River Plant.
Esso Resources Canada Ltd.	Goodlow	0.65	-		-	Sales gas to Westcoast Transmission Line. Acid gas is flared.
Gas Trunk Line of B.C. Ltd.	Boundary Lake	0.56				1983 est.
Gas Trunk Line of B.C. Ltd.	Kobes Creek	4.24	-		-	Sent to Pine River Plant. Acid gas is flared.
Northern Utilities B.C. Ltd.**	Tumbler Ridge	0.26	0.23		0.18	For the town of Tumbler Ridge
Quasar Petroleum Ltd.	Grizzly Valley	2.46	-		-	Sent to Pine River Plant

*including LPG **Proposed

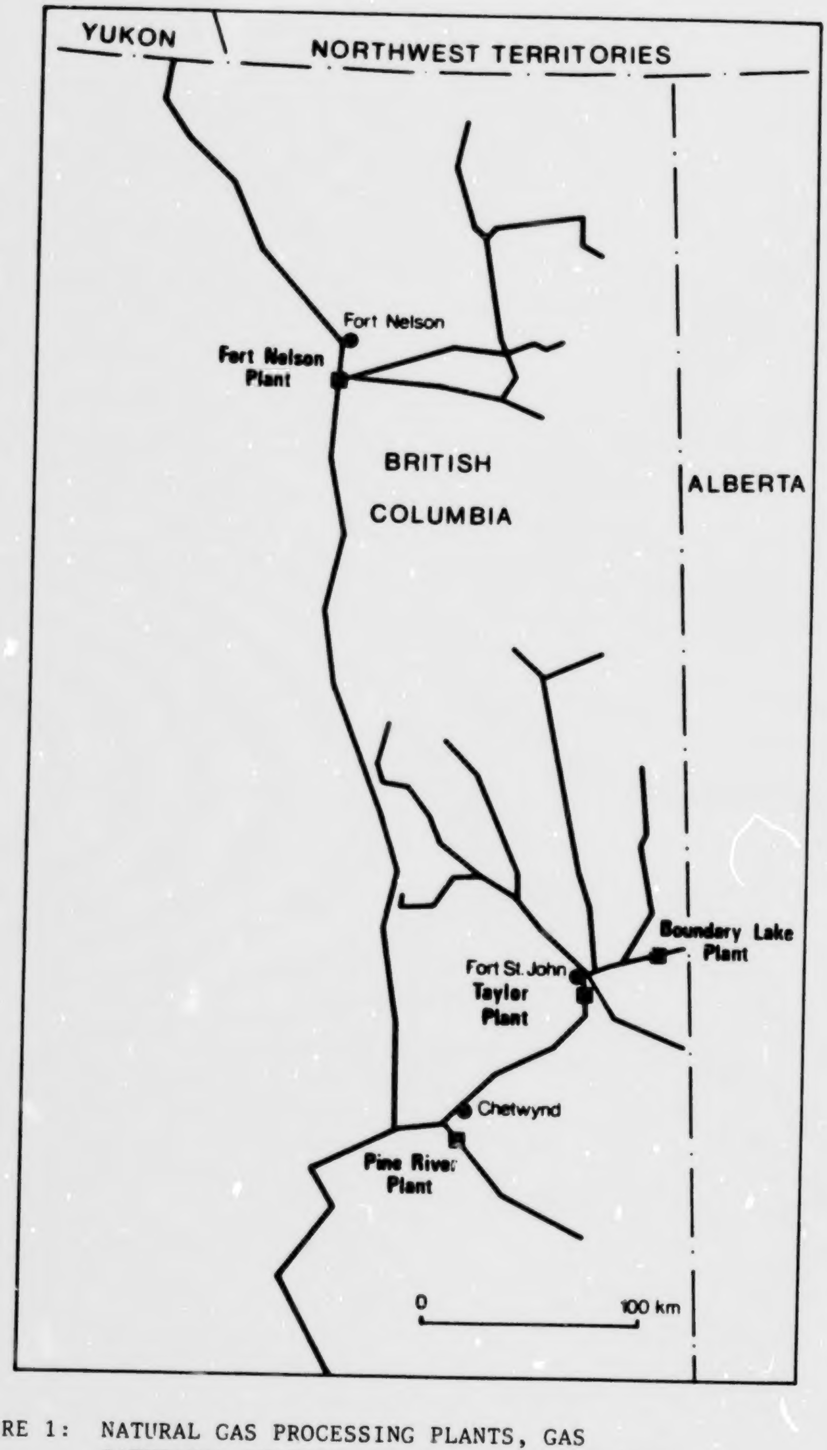

YUKON
NORTHWEST TERRITORIES
Fort Nelson
Fort Nelson Plant
BRITISH COLUMBIA
ALBERTA
Boundary Lake Plant
Fort St. John
Taylor Plant
Chetwynd
Pine River Plant
0
100 km

FIGURE 1: NATURAL GAS PROCESSING PLANTS, GAS GATHERING SYSTEMS AND MAJOR TRANSMISSION PIPELINES IN NORTHEASTERN BRITISH COLUMBIA.

SOURCE: (2)

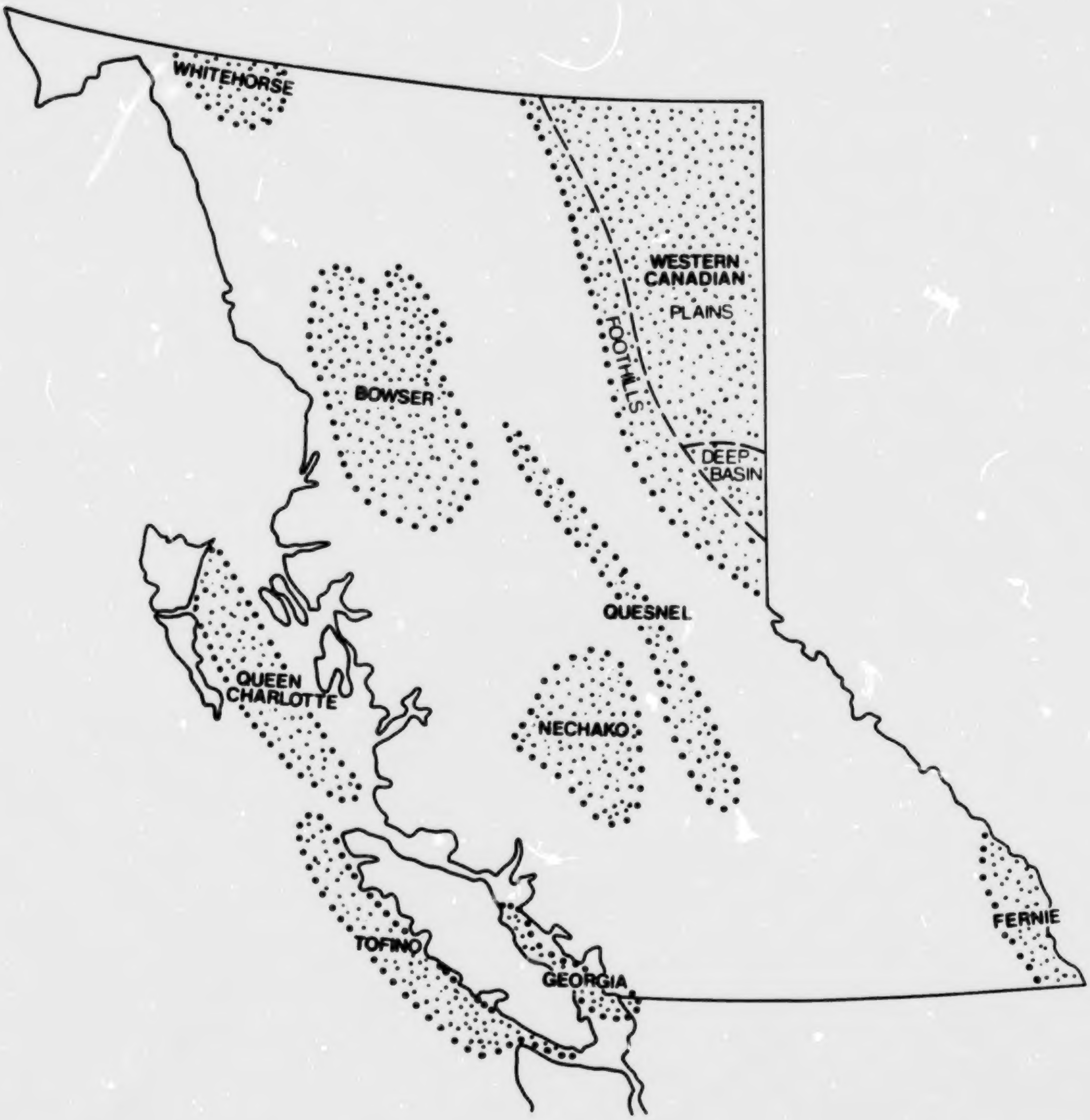

WHITEHORSE
WESTERN CANADIAN
PLAINS
FOOTHILLS
DEEP BASIN
BOWSER
QUESNEL
QUEEN CHARLOTTE
NECHAKO
TOFINO
GEORGIA
FERNIE

FIGURE 2: LOCATIONS OF MAJOR SEDIMENTARY BASINS IN BRITISH COLUMBIA

SOURCE: (2)

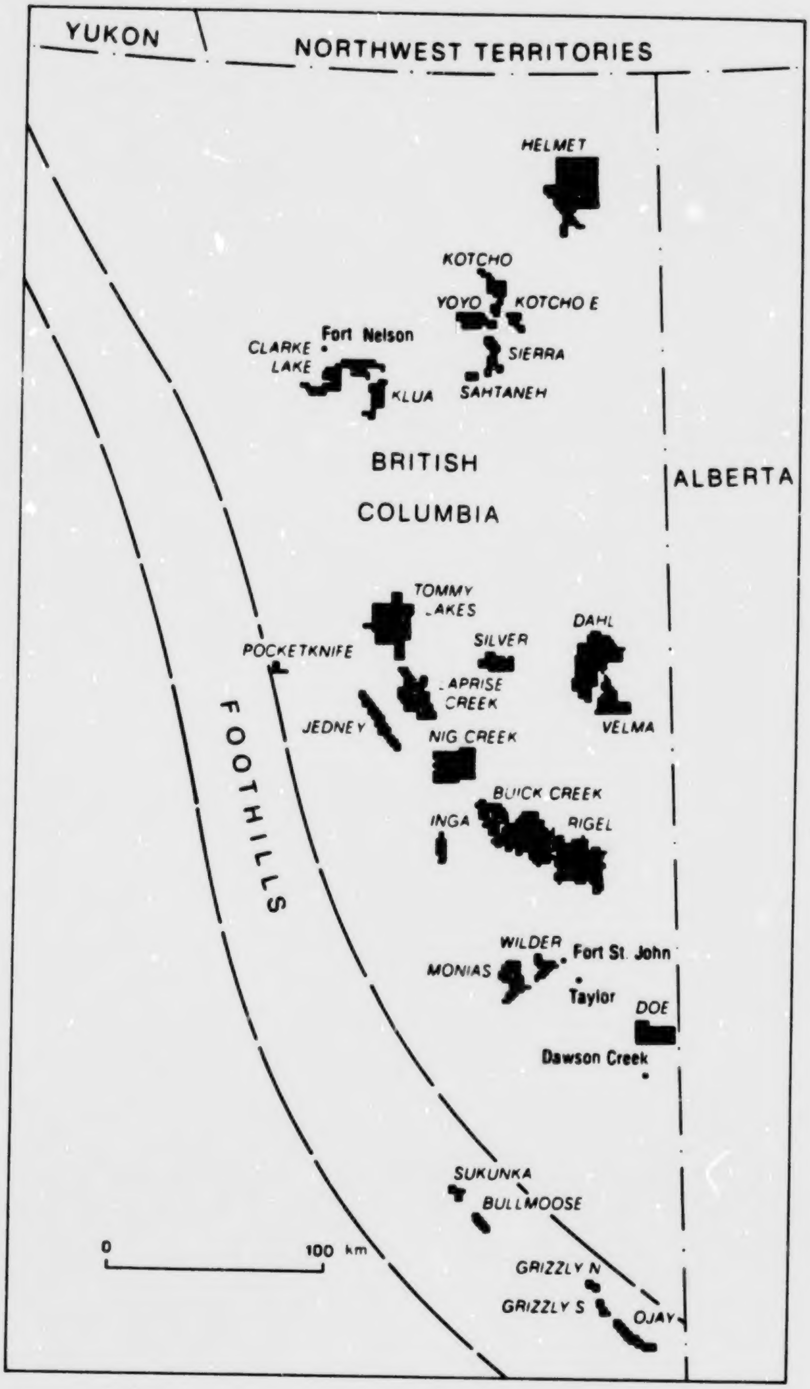

YUKON
NORTHWEST TERRITORIES
HELMET
KOTCHO
YOYO
KOTCHO E
Fort Nelson
CLARKE LAKE
SIERRA
KLUA
SAHTANEH
BRITISH COLUMBIA
ALBERTA
TOMMY LAKES
DAHL
SILVER
POCKETKNIFE
LAPRISE CREEK
JEDNEY
VELMA
NIG CREEK
FOOTHILLS
BUICK CREEK
INGA
RIGEL
WILDER
Fort St. John
MONIAS
Taylor
DOE
Dawson Creek
SUKUNKA
BULLMOOSE
0
100 km
GRIZZLY N
GRIZZLY S
OJAY

FIGURE 3: LOCATIONS OF MAJOR GAS FIELDS IN THE WESTERN CANADIAN BASIN IN NORTHEASTERN BRITISH COLUMBIA

SOURCE: (2)

TABLE 2

ESTIMATES OF REMAINING UNDISCOVERED AND ULTIMATE POTENTIAL OF MARKETABLE NATURAL GAS IN BRITISH COLUMBIA

(x 10^9 cubic metres)

Area	Initial Established Reserves*	Remaining Undiscovered Potential	Ultimate Potential
Northeast			
Foothills	48	230	278
Plains Paleozoic	149	170	319
Plains Mesozoic	191	60	251
Deep Basin	4	40	44
Subtotal (rounded)	400	500	900
Queen Charlotte and Tofino	0	100	100
Other Basins	0	Not evaluated	
Total province	400	600	1 000

*Estimates as of December 31, 1981 by the Ministry of Energy, Mines and Petroleum Resources

Source: (2)

The "pentane+" group includes varying amounts of higher hydrocarbons like pentane, hexane, heptane and octane, and is also referred to as naphtha or natural gasoline.

Besides the hydrocarbons and water vapour, other undesirable compounds often present in the reservoir gas are carbon dioxide and sulphurous gases. Carbon dioxide does not contribute to the heating value of the natural gas, and acts as a diluent only. Although H_2S is usually the major constituent of the total sulphurous gases present in the reservoir gas, others which may be found in minor quantities are CS_2, COS, SO_2 and RSH. Hydrogen sulphide is toxic and in presence of moisture corrosive too.

The raw gas composition varies from one gas field to another, even from one well to another in the same field. The proportions of impurities mentioned above may vary from zero to a high percentage. As the composition of the sales gas, typical analysis of which is shown in Table 3, is virtually constant, the extent of processing required would depend on the characteristics of the particular raw gas supply. The gas composition from a well, however, changes with time as the gas is gradually withdrawn.

5. NATURAL GAS PROCESSING(4-7)

For both economic and technical reasons, it is a common practice to gather raw gas from various fields and to process the mixed gas in a central plant. In a majority of the cases, however, some preliminary treatment of the reservoir gas at the well-head becomes necessary prior to its transmission to the central plant. The extent of such treatment of the well-head gas would depend on several factors such as the quality of the reservoir gas, the location of the well relative to the central processing plant, availability of the utilities, etc.

TABLE 3

TYPICAL ANALYSIS OF SALES GAS QUALITY NATURAL GAS

Constituent	Mole %
Methane	91.57
Ethane	5.01
Propane	0.25
N-butane	Trace
Iso-butane	0.01
N-pentane	Trace
Iso-pentane	Trace
Hexane$^+$	Nil
Nitrogen	2.65
Carbon dioxide	0.49
Oxygen	Nil
Hydrogen	Nil
TOTAL:	100.00

Total sulphur, including 2.75 mg/m^3 of odourant, content is 5.26 mg/m^3.

Source: (3)

However, there are several common unit processes applied for gas treatment at the well-head as well as the central plant. Slight variations of these processes may be practised at the well-head because of both technical and economic reasons, although the purpose and principle remain the same. For the sake of brevity the key features of the basic processes employed at both the well-head and the central plant are discussed here.

As an aid to the understanding of the basic steps therein, various gaseous stream compositions are shown in Table 4.

The major processes involved in transforming reservoir gas to the sales gas specifications could be listed as:

1. Free Water Removal,
2. Gas Separation,
3. Gas Purification,
4. Gas Conditioning,
5. Auxiliary Processes,
6. By-product Recovery, and
7. Pollution Control.

Depending on the circumstances some or all of these processes may be required at the well-head and the central plant. The individual processes are described briefly here with the exception of pollution control processes which are dealt with in subsequent sections.

Typical alternative well-head installations are shown schematically in Figures 4(a)-(d). Minimum treatment of the reservoir gas is necessary when the well is located close to the central processing plant and the gas is relatively dry and sweet (Figure 4(a)). However, if the

TABLE 4

EXAMPLES OF COMPOSITION OF VARIOUS GASEOUS STREAMS

Stream	Mole % (dry basis)													
	CH_4	C_2H_6	C_3H_8	iC_4H_{10}	nC_4H_{10}	C_5+	H_2S	CS_2	COS	SO_2	CO_2	N_2	O_2	$S_{vap.}$	Total
Raw Gas	66.15	3.78	1.05	0.20	0.78	6.05	16.84				4.15	1.00			100.00
Sour Gas	73.58	3.65	0.83	0.12	0.27	0.13	16.02				4.28	1.12			100.00
Sales Gas	91.90	5.02	0.35	Trace	Trace	Trace					0.43	2.30			100.00
Acid Gas	1.27	0.19	0.07	0.03	0.05		80.12				18.24	0.03			100.00
Tail Gas							1.92	0.40	0.59	0.72	11.82	84.55		Trace	100.00
Sour Condensates	10.15	3.10	3.02	0.49	4.22	50.24	26.73				2.05	Trace			100.00
Propane		3.00	96.12	0.88	Trace										100.00
Butane			0.86	32.16	66.08	0.90									100.00
Pentanes+			0.03	0.35	3.43	96.16	0.03								100.00
Tail Gas Incinerator Stack Gas										1.50	7.55	87.30	3.65		100.00

Note: The above analyses do not represent a mass balance of any processing plant.

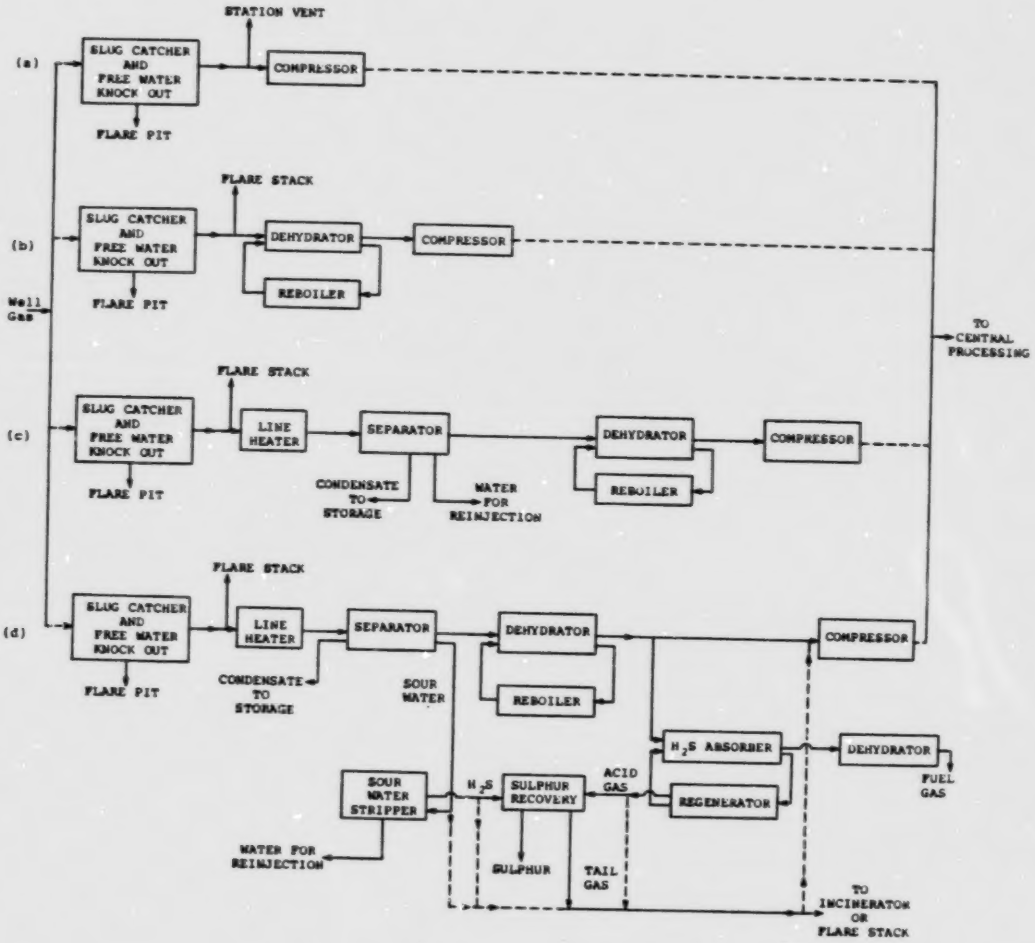

STATION VENT
(a)
SLUG CATCHER AND FREE WATER KNOCK OUT
COMPRESSOR
FLARE PIT
FLARE STACK
(b)
SLUG CATCHER AND FREE WATER KNOCK OUT
DEHYDRATOR
COMPRESSOR
REBOILER
FLARE PIT
Well Gas
TO CENTRAL PROCESSING
FLARE STACK
(c)
SLUG CATCHER AND FREE WATER KNOCK OUT
LINE HEATER
SEPARATOR
DEHYDRATOR
COMPRESSOR
REBOILER
FLARE PIT
CONDENSATE TO STORAGE
WATER FOR REINJECTION
FLARE STACK
(d)
SLUG CATCHER AND FREE WATER KNOCK OUT
LINE HEATER
SEPARATOR
DEHYDRATOR
COMPRESSOR
FLARE PIT
CONDENSATE TO STORAGE
SOUR WATER
REBOILER
H2S ABSORBER
DEHYDRATOR
FUEL GAS
SOUR WATER STRIPPER
H2S
SULPHUR RECOVERY
ACID GAS
REGENERATOR
WATER FOR REINJECTION
SULPHUR
TAIL GAS
TO INCINERATOR OR FLARE STACK

FIGURE 4: SCHEMATICS OF ALTERNATIVE WELL-HEAD FACILITIES

reservoir gas is really sour and no sweet fuel gas is available at the site, a small-scale version of the main processing plant is required at the well-head, as shown in Figure 4(d). A typical flow diagram of the central gas processing plant is shown in Figure 5.

5.1 Free Water Removal

The presence of free water in the reservoir gas is a potential cause of trouble in equipment downstream, and it is usually separated in an entrainment-type separator. The free water knockout drum also captures substances like drilling mud, corrosion inhibitors, formation solids, etc. present in the gas. On occasions and in cold climate operations, the presence of water in the well gas causes formation of hydrates in the pipes and other equipment. In such cases the raw gas is heated in a line heater just above the hydrate formation temperature and then passes through either another separator or a dehydrator. Line heaters are usually of indirect type heat exchangers in which a hot fluid passing through coils transmits the required heat to the surrounding gas. Direct fired heaters may also be used, but these generally require close supervision.

5.2 Gas Separation

Under reservoir conditions of high pressure, the liquid substances remain mixed with the gas as vapours. These liquids are hydrocarbon condensates and water which require separation before the treatment of the gas to produce sales gas. On occasions it becomes necessary to remove these components, at least partially, from the well gas at the well-head to prevent any transmission problem. Further separation of liquids from the mixed gas is

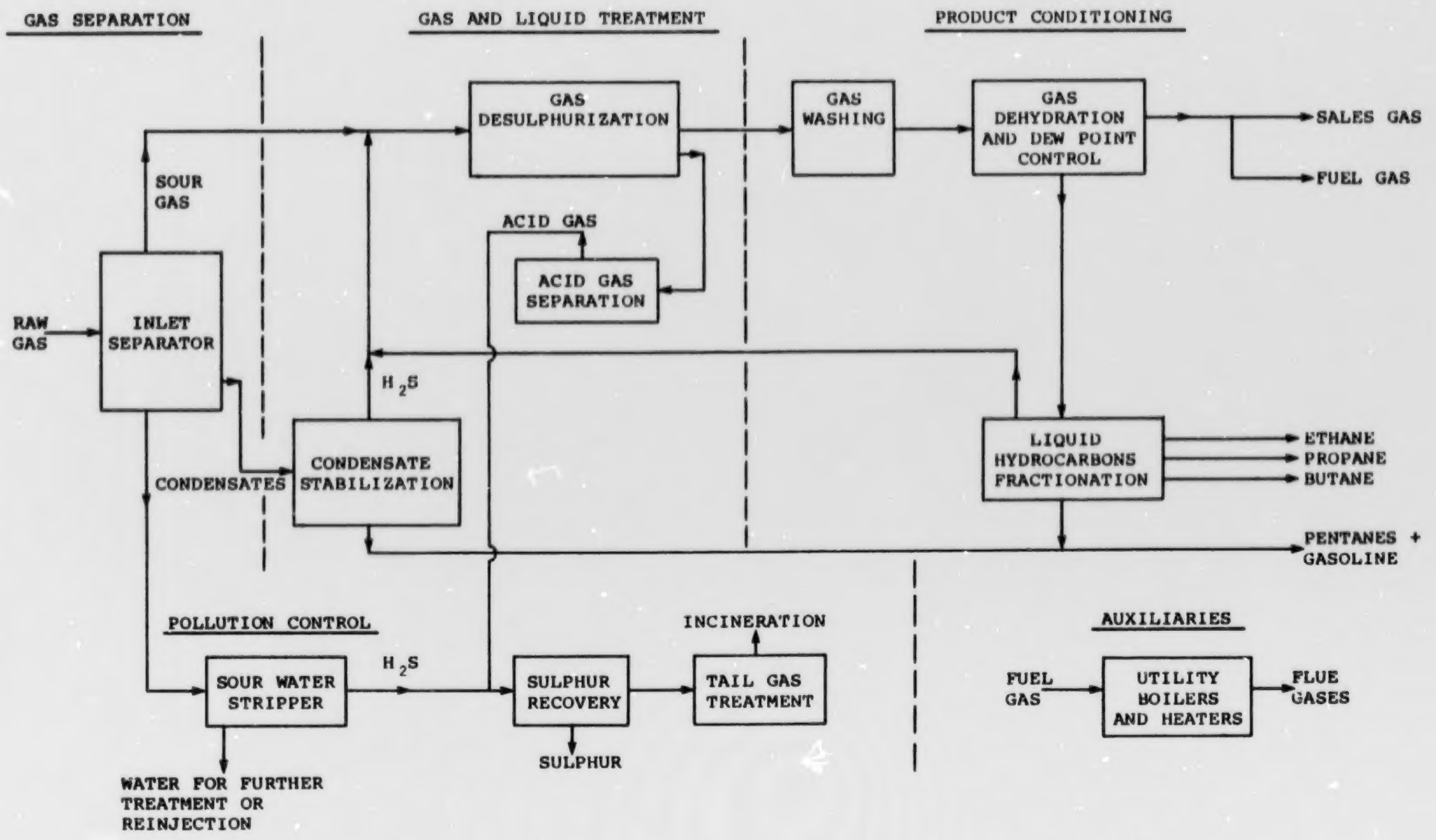

GAS SEPARATION
GAS AND LIQUID TREATMENT
PRODUCT CONDITIONING
RAW GAS
INLET SEPARATOR
SOUR GAS
CONDENSATES
CONDENSATE STABILIZATION
H_2S
GAS DESULPHURIZATION
ACID GAS
ACID GAS SEPARATION
GAS WASHING
GAS DEHYDRATION AND DEW POINT CONTROL
SALES GAS
FUEL GAS
LIQUID HYDROCARBONS FRACTIONATION
ETHANE
PROPANE
BUTANE
PENTANES + GASOLINE
POLLUTION CONTROL
SOUR WATER STRIPPER
WATER FOR FURTHER TREATMENT OR REINJECTION
H_2S
SULPHUR RECOVERY
SULPHUR
TAIL GAS TREATMENT
INCINERATION
AUXILIARIES
FUEL GAS
UTILITY BOILERS AND HEATERS
FLUE GASES

FIGURE 5: TYPICAL FLOW DIAGRAM OF A GAS PROCESSING PLANT

usually carried out as a first stage of processing at the central plant.

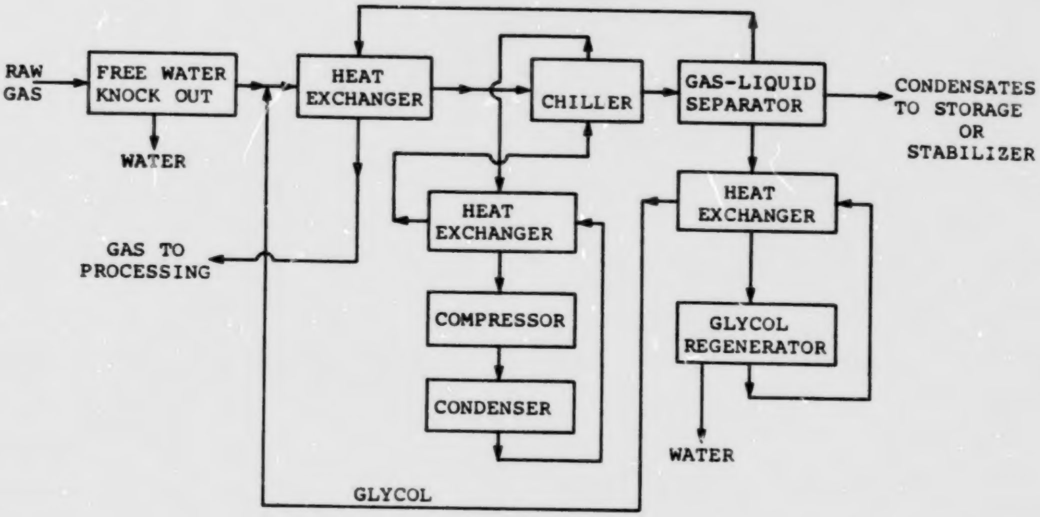
The separation of liquids from the gas can be carried out by gravity separation, absorption, adsorption, refrigeration or a combination of any of these methods. At a well-head installation it is sometimes advantageous to practice liquid separation and gas dehydration simultaneously. Typical flow diagrams of two such systems are shown in Figure 6. The most common absorbent and adsorbent used in such installations are glycol solution and molecular sieve or silica gel respectively.

In a central plant the mixed raw gas is passed through an inlet separator, which may be a vertical, horizontal or spherical vessel. The vertical type is considered to be the most versatile and widely used. The separation is normally carried out by providing centrifugal force and sufficient retention time for the inlet gas inside the vessel. While the separated gas from the top of the separator is sent for further treatment, water and condensates are drained from the bottom of the vessel. A typical vertical inlet separator is shown schematically in Figure 7.

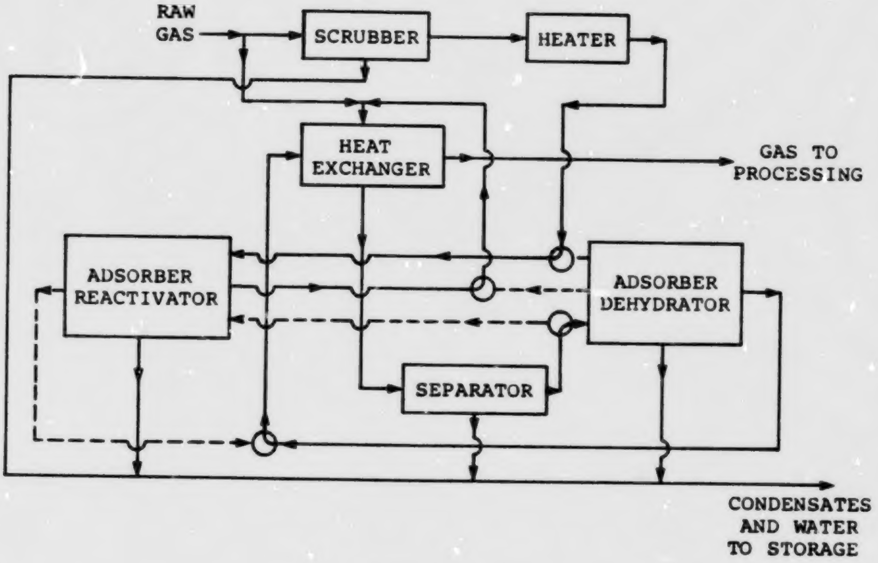
Condensates are usually processed further in a stabilizer for recovery of by-products, and the water is treated for pollution control, if necessary, before re-injection into the well or disposal by other means. These are discussed in later sections.

5.3 Gas Purification(4-20)

This is considered to be the most important part of the whole gas processing system, and mainly consists of desulphurization of the sour gas. A number of processes,


RAW GAS
FREE WATER KNOCK OUT
WATER
HEAT EXCHANGER
CHILLER
GAS-LIQUID SEPARATOR
CONDENSATES TO STORAGE OR STABILIZER
HEAT EXCHANGER
HEAT EXCHANGER
GAS TO PROCESSING
COMPRESSOR
CONDENSER
GLYCOL REGENERATOR
WATER
GLYCOL

REFRIGERATION-ABSORPTION PROCESS


RAW GAS
SCRUBBER
HEATER
HEAT EXCHANGER
GAS TO PROCESSING
ADSORBER REACTIVATOR
ADSORBER DEHYDRATOR
SEPARATOR
CONDENSATES AND WATER TO STORAGE

DRY DESICCANT ADSORPTION PROCESS

FIGURE 6: TYPICAL FLOW DIAGRAMS OF GAS-LIQUID SEPARATION SYSTEMS

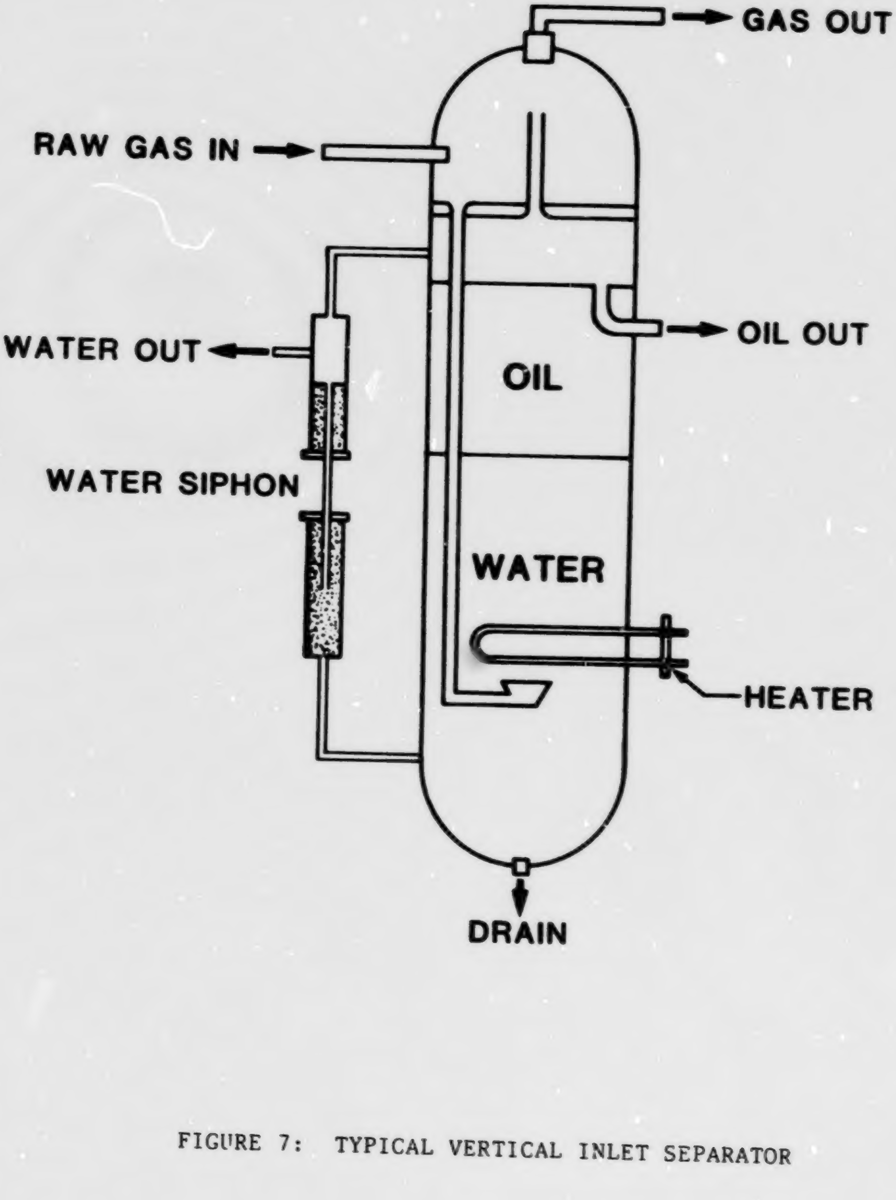

GAS OUT
RAW GAS IN
OIL OUT
WATER OUT
OIL
WATER SIPHON
WATER
HEATER
DRAIN

FIGURE 7: TYPICAL VERTICAL INLET SEPARATOR

as listed in Table 5, are available for removal of sulphur compounds from the sour gas, and the subject matter could be the topic of a separate report. The choice of a process would primarily depend on the sour gas quality, the degree of desulphurization required and the economics of the process. In British Columbia the existing gas processing plants use mainly an alkanolamine solution to absorb sulphur compounds from the raw gas (Table 6). Most commonly used solutions are monoethanolamine (MEA), diethanolamine (DEA) and Sulfinol, a mixture of di-isopropanolamine (DIPA) and sulpholane in water. At Fort Nelson, part of the sour gas is treated with hot potassium carbonate solution in conjunction with MEA. Where no sweet gas is available as fuel, it may sometimes be necessary to sweeten a portion of the well gas for use at the well-head. Most of such facilities have MEA or DEA absorbers, and others use amine and glycol solution in water, amine cartridges or molecular sieve columns. The key features of the processes used in the plants in British Columbia are listed in Table 7.

Typical flow diagrams of the above desulphurization processes are shown schematically in Figures 8(a)-(c). The sour gas from the inlet separator enters at the bottom of the absorption tower and moves upwards counter-current to the flow of the absorbent solution which is fed from the top. The strength of the absorbent solution varies from about 15 to 80%, depending on the H_2S content of the inlet gas and the solvent composition. On absorption unstable soluble salts are formed by reaction between the sulphur species and the absorbent. The wet sweet gas comes out at the top of the absorber, and is sent to the conditioning section. The spent rich absorbent is discharged into a flash tank and then to a regenerator.

TABLE 5

SOUR GAS DESULPHURIZATION PROCESSES

I Wet-bed Process				II	III
Chemical-solvent Process					
Alkanolamines	Alkaline salts	Physical-solvent Process	Special-solvent Process	Direct Conversion Process	Dry-bed Process
DEA DGA MDEA MEA SNPA-DEA TEA	Alkacid Benfield Catacarb Giammarco-Vetrocoke Hot Pot Seaboard Vacuum Carbonate	Estasolvan Fluor solvent Purisol Rectisol Selexol Sepasolh MPE Sulfinol	Ammonex Lacy-Keller M-pyrol Nalco	Ferrox Giammarco-Vetrocoke Haines IFP Lacy-Keller Mahchester Nalco Perox Shell Molecular sieves Stretford Takahax Thylox Townsend	Haines Iron Sponge Shell Molecular sieves

DEA - diethanolamine
DGA - diglycolamine (process of Fluor Corp., also called Econamine)
IFP - Institut Francais de Petrole
MDEA - methyldiethanolamine
MEA - monoethanolamine
SNPA-DEA - Societe Nationale des Petroles d'Aquitaine (SNPA) process using DEA
TEA - triethanolamine
Sepasolv MPE - special mixture of oligoethylene glycol methyl isopropyl ethers

TABLE 6

SOUR GAS DESULPHURIZATION PROCESSES USED IN BRITISH COLUMBIA

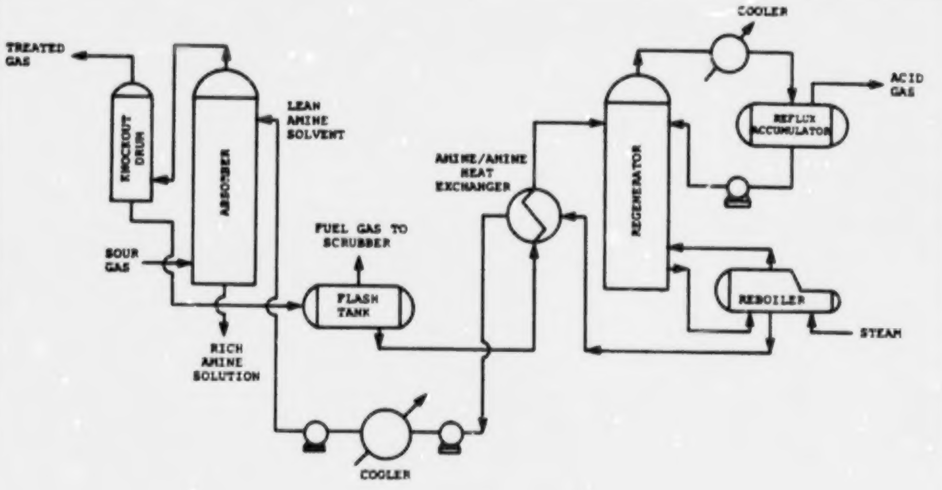
Plant and Location	Process	Remarks
Major Plants		
Petro-Canada Exploration Inc. and Westcoast Transmission Co. Ltd., Taylor	MEA solvent	Five processing trains
Westcoast Transmission Co. Ltd., Fort Nelson	Hot potassium carbonate, MEA and DEA solvents	Four processing trains using hot potassium carbonate and MEA; four processing trains using DEA.
Westcoast Transmission Co. Ltd., Pine River	Sulfinol solvent	Two processing trains
Other Plants		
Amoco Canada Petroleum Co. Ltd., Beaver River	Molecular sieve column	
B.P. Exploration Canada Ltd., Bullmoose	DEA	
B.P. Exploration Canada Ltd., Sukunka	DEA	
Esso Resources Canada Ltd., Goodlow	MEA solvent	
Gas Trunk Line of B.C. Ltd., Boundary Lake	MEA solvent	One processing train
Gas Trunk Line of B.C. Ltd., Kobes Creek	MEA or DEA	
Northern Utilities B.C. Ltd., Tumbler Ridge	DEA	One processing train with an emergency iron sponge absorber

A train consists of individual process vessels, such as absorber column, regenerator column, reboiler, coolers, pumps, piping system, etc.

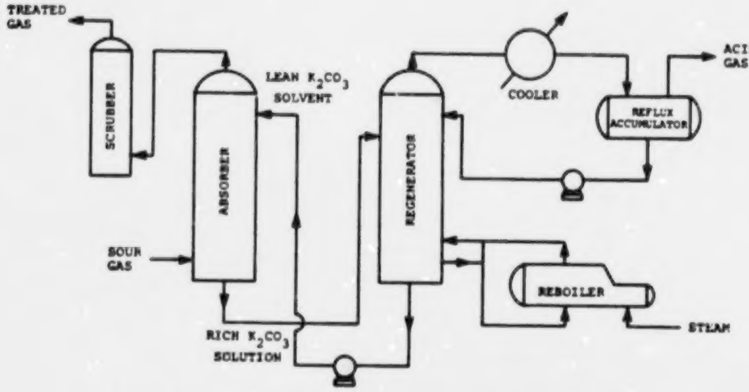
TABLE 7

MAJOR FEATURES OF SOUR GAS DESULPHURIZATION PROCESSES IN USE IN BRITISH COLUMBIA

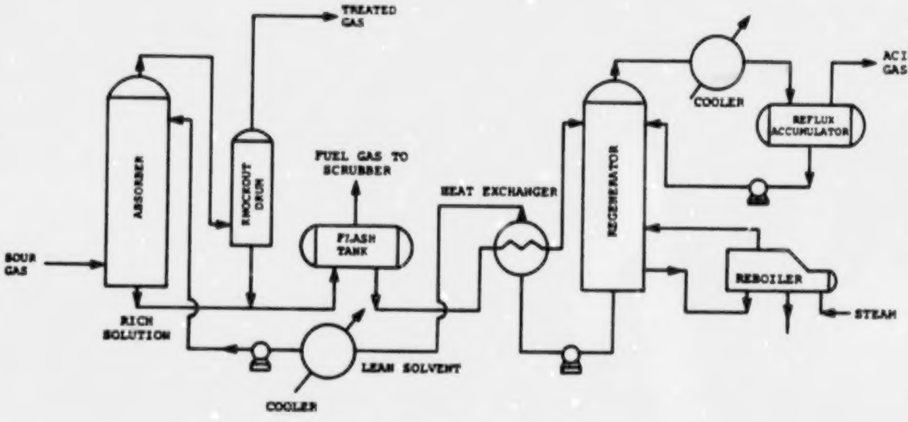
Process	Feed Gas Characteristics	Process Steps	Remarks
MEA	High volume of sour gas containing low to medium H_2S concentration.	Absorption of H_2S and solvent regeneration by heat stripping.	High reactivity towards H_2S and CO_2; inability to remove RSHs and to cope with COS and CS_2.
DEA	High volume of sour gas containing high H_2S concentration.	Absorption of acid components and solvent regeneration by heat stripping.	Less reactive than MEA but resistant to degradation by COS and CS_2, and lower vaporization losses than MEA.
Hot Potassium Carbonate	High volume of sour gas with appreciable CO_2 content to provide low H_2S/CO_2 ratio.	Absorption of acid components under high pressure, regeneration of solvent by pressure reduction and heat stripping.	Effectively removes COS and CS_2, but RSHs are not readily absorbed. Solvent is stable and has low vapourization losses.Usually an amine unit is required as a follow-up.
Sulfinol	High volume of sour gas with low to high acid gas partial pressure and H_2S/CO_2 ratio 1:1 or greater.	Absorption of acid components by both chemical and physical solvent actions. Solvent regeneration by heat stripping.	Effectively removes COS, CS_2 and RSHs; CO_2 removal could be pre-determined. Low solvent degradation rate and low vapourization losses. Solvent composition can be varied to suit particular application.


TREATED GAS
KNOCKOUT DRUM
ABSORBER
LEAN AMINE SOLVENT
SOUR GAS
RICH AMINE SOLUTION
FUEL GAS TO SCRUBBER
FLASH TANK
AMINE/AMINE HEAT EXCHANGER
COOLER
REGENERATOR
COOLER
REFLUX ACCUMULATOR
ACID GAS
REBOILER
STEAM

(a) AMINE PROCESS


TREATED GAS
SCRUBBER
ABSORBER
LEAN K_2CO_3 SOLVENT
SOUR GAS
RICH K_2CO_3 SOLUTION
REGENERATOR
COOLER
REFLUX ACCUMULATOR
ACID GAS
REBOILER
STEAM

(b) HOT POTASSIUM CARBONATE PROCESS


TREATED GAS
ABSORBER
KNOCKOUT DRUM
SOUR GAS
RICH SOLUTION
FUEL GAS TO SCRUBBER
FLASH TANK
HEAT EXCHANGER
LEAN SOLVENT
COOLER
REGENERATOR
COOLER
REFLUX ACCUMULATOR
ACID GAS
REBOILER
STEAM

(c) SULFINOL PROCESS

FIGURE 8: TYPICAL FLOW SCHEMATICS OF DESULPHURIZATION SYSTEMS

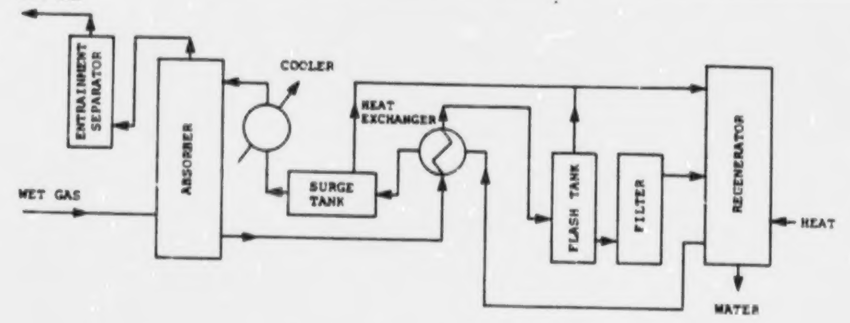
In the regenerator the spent absorbent is heated at a reduced pressure as well as refluxed to strip off the acid gases. The regenerated lean solvent is cooled and recycled to the absorption tower. The acid gases are either flared or processed for sulphur recovery depending on the regulatory requirements.

5.4 Gas Conditioning

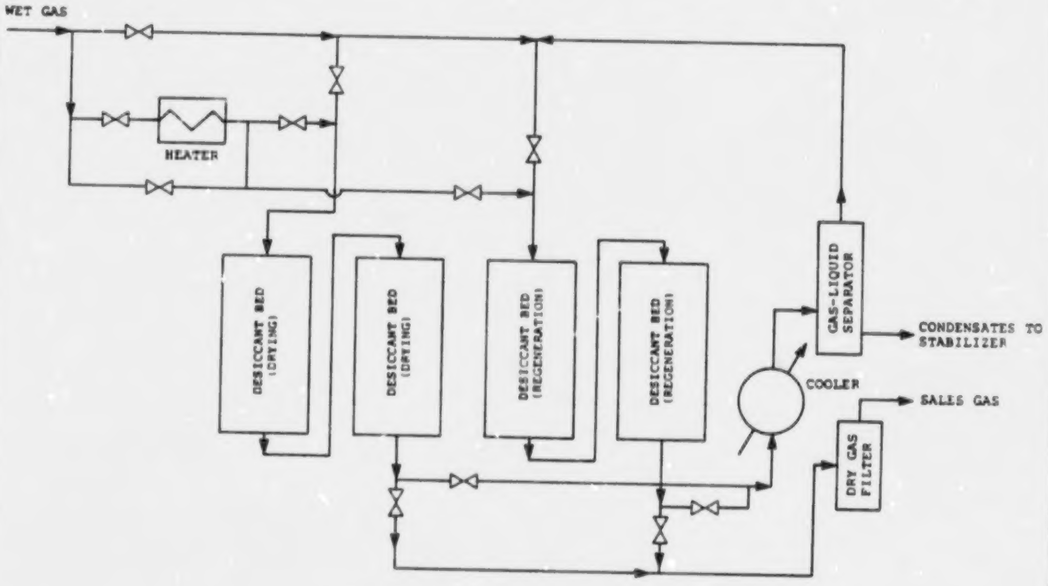
The desulphurized gas from the purification system is given final conditioning to meet water and hydrocarbon dew point specifications of the sales gas. At this stage of processing the gas is first cooled and washed with water to remove traces of absorbent carried over from the desulphurization unit. The washing is usually carried out in a column with water being sprayed from the top while the gas flows upwards counter-currently from the bottom.

The wet gas is finally dehydrated, and any liquid hydrocarbons still present in the gas are recovered at this stage. Both water and liquid hydrocarbons could be removed by absorption, adsorption or refrigeration or a combination of any of these processes. Each method has its merits and demerits, so the most cost-effective system is chosen for each case. Brief descriptions of these processes are given below and their typical flow diagrams are shown in Figures 9(a)-(c).

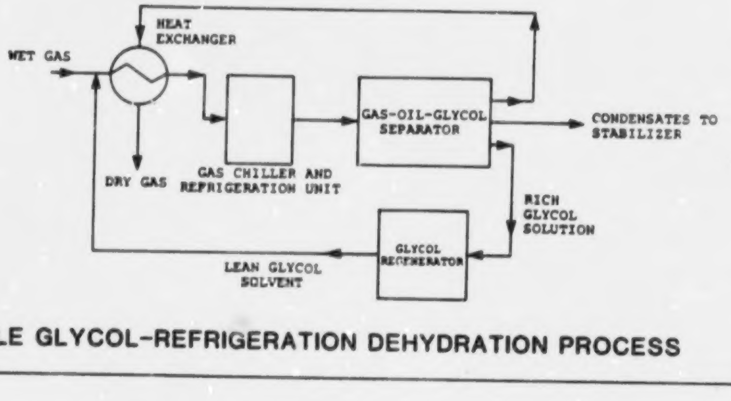
In a glycol dehydrator system (Figure 9(a)) the wet gas is passed counter-currently through an absorber. The absorber is usually a trayed column where the gas is fed at the bottom and lean glycol solution flows from the top. Normally triethylene glycol (TEG) is used as the absorbent, although diethylene glycol (DEG) and glycol mixtures are also employed in some facilities. The lean glycol absorbs the water from the gas, and the dry sales


DRY GAS
ENTRAINMENT SEPARATOR
ABSORBER
COOLER
HEAT EXCHANGER
SURGE TANK
WET GAS
FLASH TANK
FILTER
REGENERATOR
HEAT
WATER

(a) GLYCOL PROCESS


WET GAS
HEATER
DESICCANT BED (DRYING)
DESICCANT BED (DRYING)
DESICCANT BED (REGENERATION)
DESICCANT BED (REGENERATION)
GAS-LIQUID SEPARATOR
CONDENSATES TO STABILIZER
COOLER
SALES GAS
DRY GAS FILTER

(b) DRY BED PROCESS


HEAT EXCHANGER
WET GAS
GAS-OIL-GLYCOL SEPARATOR
CONDENSATES TO STABILIZER
DRY GAS
GAS CHILLER AND REFRIGERATION UNIT
RICH GLYCOL SOLUTION
GLYCOL REGENERATOR
LEAN GLYCOL SOLVENT

(c) SIMPLE GLYCOL-REFRIGERATION DEHYDRATION PROCESS

FIGURE 9: TYPICAL FLOW SCHEMATICS OF DEHYDRATION SYSTEMS

gas is passed through an entrainment separator to remove traces of the absorbent. The spent glycol solution is regenerated by heating. While the lean glycol is cooled and recycled to the absorber, the water removed is usually vented to the atmosphere. The major factors affecting the dehydration efficiency are concentration of glycol in the lean solution, glycol circulation rate and absorber contact time.

A dry-bed desiccant system (Figure 9(b)) uses solid adsorbents, such as silica gel or molecular sieve, to remove water and liquid hydrocarbons from the wet gas. It normally consists of two or more adsorbent columns arranged in parallel and which operate on a predetermined timed cycle. While one column is in use for dryng the wet gas, a second one is regenerated by hot gas at the same time. The gas flows are switched from one column to another at the end of each timed cycle. The gases from the regenerator are cooled, and water and hydrocarbons are separated. The dry gas from the adsorbing column is filtered and sent to the pipeline.

In a simple refrigeration plant (Figure 9(c)), the gas is cooled in a heat exchanger to condense out the water and heavy hydrocarbons. To prevent the formation of hydrates during cooling, normally glycol is injected into the gas before the heat exchanger. Low-temperature separators are used to separate glycol-water, liquid hydrocarbons and dry gas. The liquid hydrocarbons are sent to either a fractionator or a stabilizer as desired. The glycol is regenerated by heating and recycled in the system.

When the recovery of "deep-cut" liquefied petroleum gases (LPG, mainly propane and butane) is required the refrigerated absorption process is normally employed. The gas from the low temperature separator in the refrigeration plant is sent through a low temperature absorber where a low molecular weight light oil is used as the absorbent. The lean oil is fed from the top of the absorber, and the gas flows upwards from the bottom. Any liquefiable hydrocarbons present in the dry gas are absorbed by the oil, which is then sent for regeneration. In this system, the rich oil is first processed in a de-ethanizer to recover absorbed methane and ethane, and then to a still where LPG and heavier hydrocarbons are removed. The lean oil from the regenerator is cooled and recycled. The glycol used upstream to prevent hydrates formation is also separated from the absorbed water and regenerated for further use in the system.

In most installations in British Columbia, the dehydration of the gas is carried out by absorption in chemicals such as DEG, TEG or a mixture of MEA and DEG in water. Solid desiccants like silica gel or molecular sieve are also used in a few installations. All these processes are regenerative and the absorbing solution or the desiccant is regenerated by heating.

5.5 Auxiliary Processes

As in any other industrial operation some supporting facilities are normally required at natural gas processing installations. Such facilities include flare stack, flare pit, boilers and heaters, compressors, turbines, and power generators for regular and/or emergency use.

The flare stack and/or the flare pit are the most essential in this particular industry for the safety of both personnel and equipment. These systems are used for the safe disposal of high pressure gas and some hydrocarbon condensates during emergencies as well as scheduled plant shutdown. Emergency situations may arise due to upsets in process equipment and gas transmission system, power failures and fires. Flare systems are generally designed to accommodate venting of either the whole supply of raw gas or partially treated gas from conveniently located areas in the treatment system.

Continuous flaring is also practised in cases where the sulphur content of the raw or acid gas is too low to warrant installation of a sulphur recovery plant. To achieve proper combustion of the flare gases and a minimum flame temperature, the usual practice in the industry is to ensure that the heat content of the gases is at least 9.32 MJ/m^3. This is often maintained by supplying supplementary fuel gas to the flare burners. A pilot flame from fuel gas burning is maintained continuously at the mouth of intermittently operated flare stacks.

In most cases elevated flare stacks are equipped with a steam injection system for smokeless burning of the gases. At any large gas processing plant there may be two flare stacks for separate flaring of low and high pressure gas streams. Some are designed with a knockout vessel to remove any entrained liquid hydrocarbons from the gas so that smokeless operation could be maintained easily. Any prolonged flaring is usually avoided and the stack height is designed for maximum dispersion of the flared gases. At some well-head installations if the reservoir gas is considered to be sweet, the pressurized gas is simply vented to the atmosphere during emergencies.

Flare pits are used to burn mainly condensates from knockout drums, compressor blowdown at well-head installations. Hydrates are also burned in these pits during "pigging" or clearing of gas supply lines. Although it is not a normal practice, sometimes small gas gathering facilities are vented through the flare pit during an emergency.

Steam generators and heaters are generally gas-fired packaged units, and used for space heating as well as in various processing equipment. Compressors and turbines are commonly used as prime movers of gas under pressure within the processing system and for pipeline transmission. Where no electricity supply is available, turbines or small generators are used for the purpose. It is also a normal practice to maintain an emergency power generator at every gas processing installation to maintain the essential equipment in operation during breakdown in power supply. The fuel burned in these units is normally sweet fuel gas or desulphurized LPG produced as by-product.

5.6 By-product Recovery

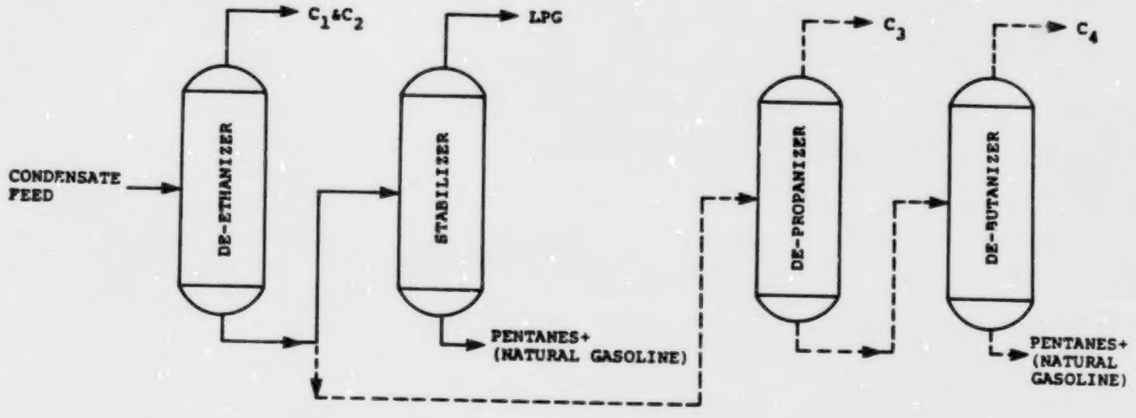
The major by-products of natural gas processing are the liquefied petroleum gases and "pentanes+". If the gas being processed is sour and environmental regulations require sulphur removal, the other by-product usually recovered is elemental sulphur. The latter recovery process is discussed under section on pollution control processes.

The recovery of hydrocarbon by-products at the well-head is seldom practised, unless any particular reservoir gas quality and the economics warrant it. The general practice is to store the condensates at the site for eventual treatment at a central gas processing plant or for sale.

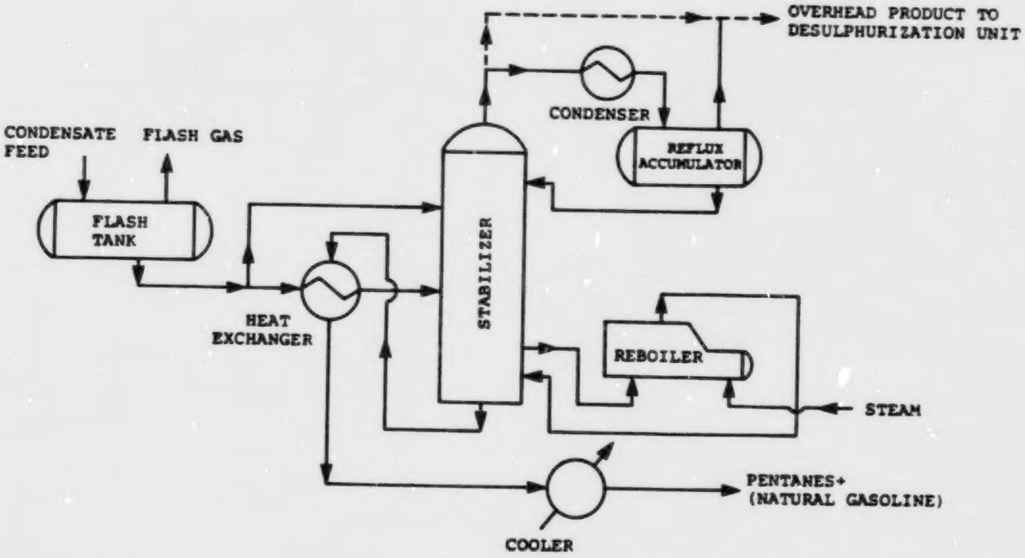
To obtain individual products of specification it is usually necessary to further process the mixed LPG and/or condensates removed from the raw gas. Depending on the characteristics of the feed liquids and the number of products to be made, the raw by-products are separated by fractionation or stabilization or both. These processes are shown schematically in Figures 10(a) and (b). In a sense both processes work under similar principles of separating products of different molecular weights by controlling operating temperature and pressure in fractionating columns. The feed enters at the middle of a column containing several trays. The operating conditions are adjusted to allow the lighter hydrocarbons to rise to the top while the heavier substances are removed from the bottom as liquids. Two or three fractionating columns may be used to produce methane, ethane, LPG (propane and butane) and pentanes plus (natural gasoline).

The condensates from the inlet separator can contain significant amounts of acid gases and lighter hydrocarbons. It is a common practice to remove these components in a flash tank by reducing the operating pressure before stabilizing the condensates in a fractionating column (Figure 10 (b)). This method of separation is economically attractive in most cases.

The fractionating column can be operated in either non-refluxed or refluxed mode. The non-refluxed system is simpler but less efficient, and suitable only for remote location. When used as a non-refluxed tower, no external cooling of the overhead product and its reflux is practised. In the refluxed mode a portion of the overhead product is condensed and recycled to the tower. The amount of reflux is controlled to achieve the maximum separation of the LPG from the feed. A reboiler is used to vapourize a portion of the bottom product which is


C_1&C_2
LPG
C_3
C_4
CONDENSATE FEED
DE-ETHANIZER
STABILIZER
DE-PROPANIZER
DE-BUTANIZER
PENTANES+ (NATURAL GASOLINE)
PENTANES+ (NATURAL GASOLINE)

(a) FRACTIONATION PROCESS


OVERHEAD PRODUCT TO DESULPHURIZATION UNIT
CONDENSER
CONDENSATE FEED
FLASH GAS
REFLUX ACCUMULATOR
FLASH TANK
STABILIZER
HEAT EXCHANGER
REBOILER
STEAM
PENTANES+ (NATURAL GASOLINE)
COOLER

(b) STABILIZATION PROCESS

FIGURE 10: TYPICAL FLOW DIAGRAMS OF CONDENSATES RECOVERY SYSTEMS

recirculated through the column to aid in the stripping of LPG from the feed. The optimum operating temperature and pressure of a fractionating column are designed on the basis of the feed composition and the products desired.

5.7 Pollution Control Processes

The major pollution control processes in a typical natural gas processing plant are acid gas treatment and combined process effluent treatment systems. Although there are other minor waste streams generated at the plant, no particular treatment is applied to any of these streams. The major processes are discussed in the following sections.

6. SOURCE, CHARACTERISTICS AND CONTROL OF WASTE STREAMS

A number of waste streams of potential environmental concern are generated in a natural gas processing plant. The major sources of these streams, the potential pollutants of concern and applicable control technologies are listed in Table 8. As the quantities of various waste discharges are site-specific, being dependent on several factors, the subject is discussed separately in Section 7.

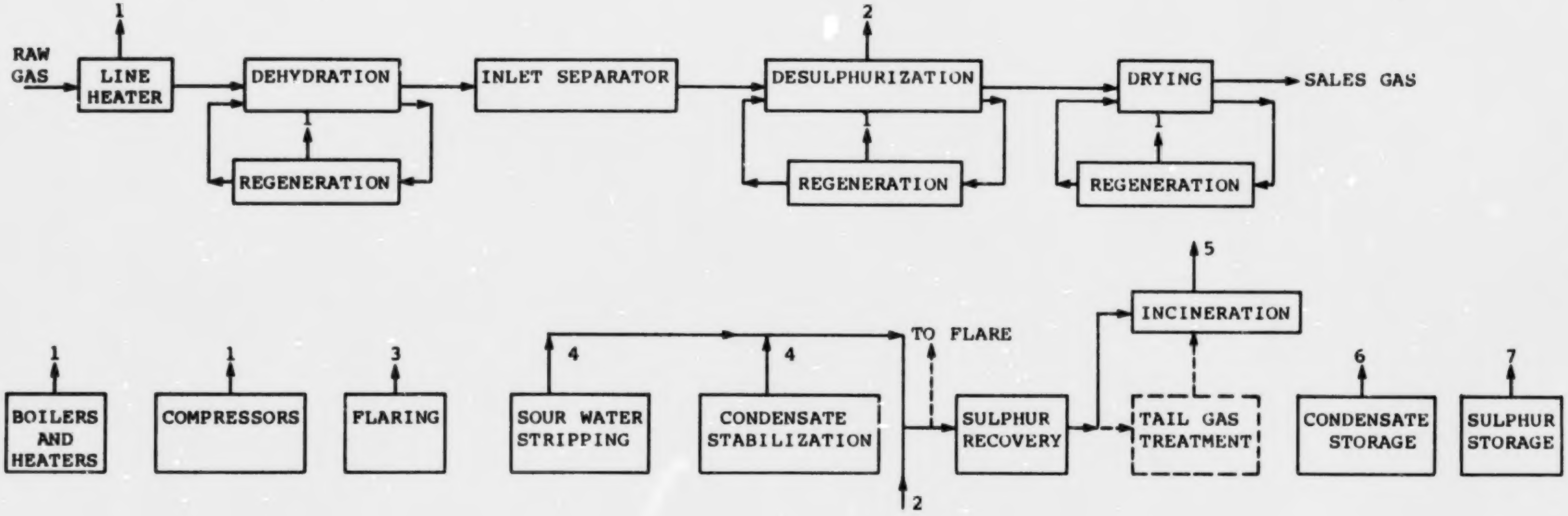
6.1 Gaseous Emissions

The sources of gaseous emission in a typical well-head installation and a central gas processing plant are shown schematically in Figure 11. The gaseous waste streams can be categorized into three groups as follows:

(i) products of combustion of fuel gas, condensates and raw gas;

TABLE 8 - MAJOR SOURCES OF WASTE GENERATION IN A GAS PROCESSING PLANT

Waste Stream or Source	Pollutants of Concern	Applicable Control Technology
Gaseous		
Line heater	SO_2, NO_x, particulates and hydrocarbons	Steam heating, desulphurized fuel gas and combustion control
Dehydrator regenerator	Water vapour, SO_2, NO_x, particulates and hydrocarbons	"
Gas sweetener regenerator	SO_2, NO_x, and hydrocarbons	"
Acid gases from gas sweetener	H_2S, COS, CS_2 and RSH	Smokeless flare, incineration, sulphur recovery and tail gas treatment
Liquid hydrocarbons recovery and condensate stabilization	H_2S, COS, CS_2, RSH and fugitive hydrocarbon vapour	Desulphurization and proper maintenance
Sour water stripping	H_2S, COS, CS_2, RSH and NH_3	"
Compressors, generators steam boilers and process heaters	SO_2, NO_x, particulates and hydrocarbons	Desulphurized fuel gas and combustion control
Station vent	Raw gas (with H_2S if sour)	Smokeless flare, incineration
Flare stack	SO_2, NO_x and smoke	Steam-injected smokeless flare
Flare pit	H_2S, SO_2, NO_x and smoke	Controlled combustion under favourable atmospheric conditions
Condensate storage	Hydrocarbon vapour	Vapour recovery
Flash tanks, pipeline valves, flanges, etc.	Fugitive emission of gases containing H_2S and hydrocarbon vapour	Vapour recovery and good maintenance
Sulphur storage and loading	Sulphur vapour or dust	Sulphur prilling and dust control measures
Liquid		
Process wastewater	Oils, hydrocarbons, chemical reagents, sulphides, phosphate, ammonia, suspended solids, COD, odour and toxicity	Oil-water separation, biological and/or physico-chemical treatment
Boiler feed water treatment and boiler blowdown	Brines, primary chlorides and sulphates of sodium, potassium, calcium and magnesium, other salts, anti-corrosive and anti-fouling chemicals, suspended solids and dissolved solids	Solids removal, recycling or combined treatment with process wastewater
Sanitary wastes	Typical sewage effluent parameters	Conventional treatment or combined treatment with process wastewater
Run-off water from sulphur storage area and other processing areas	Suspended solids and oil and grease	Oil and grease removal, solids removal, neutralization, recycling or combined treatment with process wastewater
Solid Wastes		
Oily sludge from free oil separator	Oily hydrocarbons	Oil recovery or incineration
Raw water treatment and other chemical treatment sludges	Leachable constituents	Proper landfill operation
Bio-sludges from wastewater treatment	—	Incineration or landfill


RAW GAS
1
LINE HEATER
DEHYDRATION
1
REGENERATION
INLET SEPARATOR
2
DESULPHURIZATION
1
REGENERATION
DRYING
1
REGENERATION
SALES GAS
1
BOILERS AND HEATERS
1
COMPRESSORS
3
FLARING
4
SOUR WATER STRIPPING
4
CONDENSATE STABILIZATION
TO FLARE
2
SULPHUR RECOVERY
5
INCINERATION
TAIL GAS TREATMENT
6
CONDENSATE STORAGE
7
SULPHUR STORAGE

1 - PRODUCTS OF COMBUSTION OF FUEL GAS
2 - ACID GAS, 3-PRODUCTS OF FLARING
4 - MOSTLY H_2S, 5 - INCINERATED GAS
6 - HYDROCARBON VAPOURS, 7 - SULPHUR DUST

FIGURE 11: POTENTIAL SOURCES OF GASEOUS EMISSION FROM A NATURAL GAS PROCESSING FACILITY

(ii) fugitive emission of hydrocarbon vapours, H_2S containing gas and sulphur dust; and

(iii) acid gas streams containing H_2S, COS, CS_2 and RSH.

6.1.1. Fuel Combustion and Flaring

In a gas processing plant there are two distinct groups of sources where fuel gas, condensates and raw gas are burned, namely various combustion equipment and flare systems.

6.1.1.1 Fuel Combustion

The characteristics of the products of combustion would cause minimum concern if desulphurized fuel gas is used along with proper control of combustion. Sulphur containing fuel would generate SO_2, the concentration of which would depend on the fuel sulphur content.

In some instances, the gas burning prime movers such as compressors, turbines and engines, produce significant amount of nitrogen oxides. The inherent nature of these internal combustion engines usually aids in the excessive generation of NO_x because of high combustion temperature and the use of considerable amount of excess air. Although recent modifications to the engine design have been helpful in reducing NO_x production considerably, the emission rate may still be in excess of the permissible limits. Any source control measure may be impractical because of the equipment size, the emission rate and the cost involved. In such cases, the only recourse is to design the stack height for adequate dispersion of the flue gases for complying with the applicable air quality criteria.

6.1.1.2 Flare Systems

Ground-level flare pits are used to burn mostly condensates mixed with water and residual gases from pipelines, knock-out drums and flash tanks during "pigging" operation. Because of the composition, variable energy-content and flow rates of the combustibles, there are inherent difficulties in sustaining proper combustion during the flaring. Smoky flares are not uncommon due to incomplete combustion at the time of adverse wind conditions. In addition to smoke, incomplete combustion of hydrocarbons may result in generation of pollutants such as aldehydes, ketones, mercaptans and similar species. Odours may also result from slow atmospheric oxidation of any spilled liquids in the flare pit.

There are several designs of ground-level flare systems with provisions for adequate fuel gas and air supplies to ensure complete combustion of the flared substances. Injection of steam into the combustion zone also aids in smokeless combustion. However, proper control of the amount of steam injected is essential for sustaining the flare, as an excessive amount of steam may extinguish the flame and release unburned hydrocarbons to the atmosphere. Where a flare pit is used occasionally, it is preferred that a burn is carried out whenever possible under favourable atmospheric conditions.

Good housekeeping and resource conservation practices are essential to minimize flaring at ground-level. Liquid hydrocarbons flowing to the flare pit could be routed through a holding tank, so that mostly the vapours from this tank need only to be flared. The liquids from the tank can be sent to the processing plant. If feasible, the vapours from the tank can also be recovered. At the well-head, non-combustible and difficult-to-burn materials should not be sent to the flare pit.

Elevated flares are also used in gas processing plant to dispose of gases and vapours which are not easily recoverable, and to vent the station during emergency conditions due to equipment failure. Most of these flares nowadays are designed to produce smokeless combustion. Recent investigations(21) in Alberta have shown that some conventional flares for disposal of sour gases containing 10 to 20% H_2S do not perform satisfactorily under unfavourable wind conditions. It has been observed that some residual H_2S is released into the atmosphere due to incomplete combustion, the flame blows out by high winds, and sometimes even the waste gases fail to ignite instantaneously. Moreover, most flares require supplementary fuel gas to maintain combustion of waste gases with low to medium heat content.

To rectify such problems of flaring waste gases, an incinerator has recently been developed by Western Research of Calgary, Alberta(21). The waste gases are burned within an open-ended, refractory-lined combustion tube. Wind shrouds are provided at both top and bottom ends of the tube. The combustion air is induced into the tube by the waste gases as well as by the natural draft created by hot flue gases. The igniter, pilot flames and sensors are provided at the bottom of the tube. To date three models of incinerators - flare-tip, in-stack and skid-mounted, have been developed based on the above principles. Typical design of a flare-tip incinerator is shown in Figure 12, which can be mounted in an existing flare stack. Others are used instead of flare stacks, and the skid-mounted incinerator is equipped with

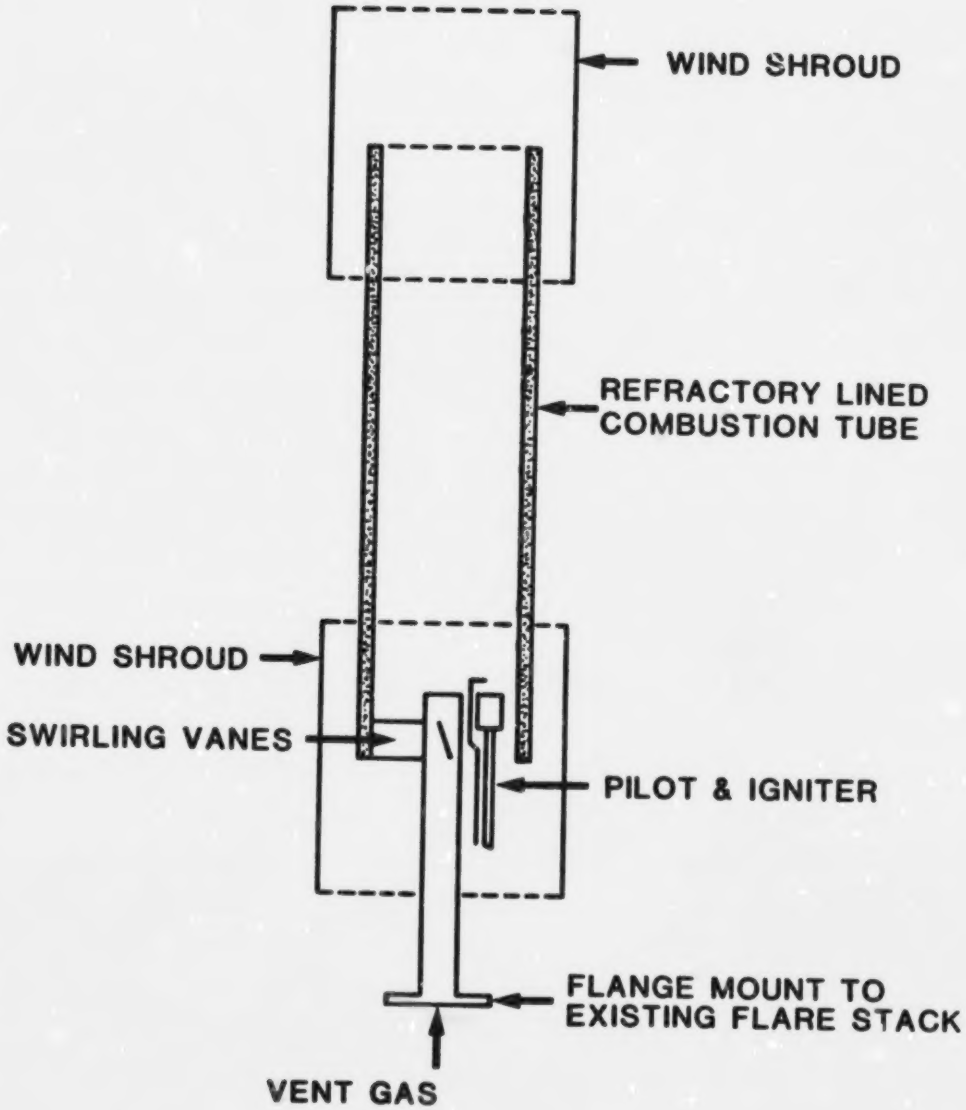

WIND SHROUD
REFRACTORY LINED COMBUSTION TUBE
WIND SHROUD
SWIRLING VANES
PILOT & IGNITER
FLANGE MOUNT TO EXISTING FLARE STACK
VENT GAS

FIGURE 12: FLARE-TIP INCINERATOR

SOURCE: (21)

a supplementary fuel supply system to burn a very low energy-content gas.

Test results confirm that an incinerator of the above kind performs better than conventional flares by ensuring complete combustion of the waste gases. It performs more reliably and unattended. The design of the incinerator allows maximum use of the available energy in the waste gases, and requires the minimum amount of supplementary fuel when it is needed. Typical test results from conventional flaring and incineration are shown schematically in Figure 13.

By the end of 1982 a total of 35 units of the above design was in operation, and a more conventional unit to incinerate 150 000 m^3/d of vent gas from several wells was under construction.

6.1.2 Fugitive Emissions

In a gas processing plant the emission of hydrocarbon vapours from storage tanks is normally not considered to be significant because of the type and volume of the stored products. Installation of vapour conservation and recovery systems in storage tanks would minimize fugitive emission from breathing and working losses. Good maintenance and housekeeping are also necessary to minimize leaks in various pumps, seals, glands, etc.

Where the liquid sulphur produced in the sulphur recovery plant is stored in pits, small amount of H_2S can be released to the atmosphere through the pit vent. Although normally it does not create a serious problem, sometimes the odour may be detectable in the vicinity. Sulphur dust deposited on the soil is transformed into

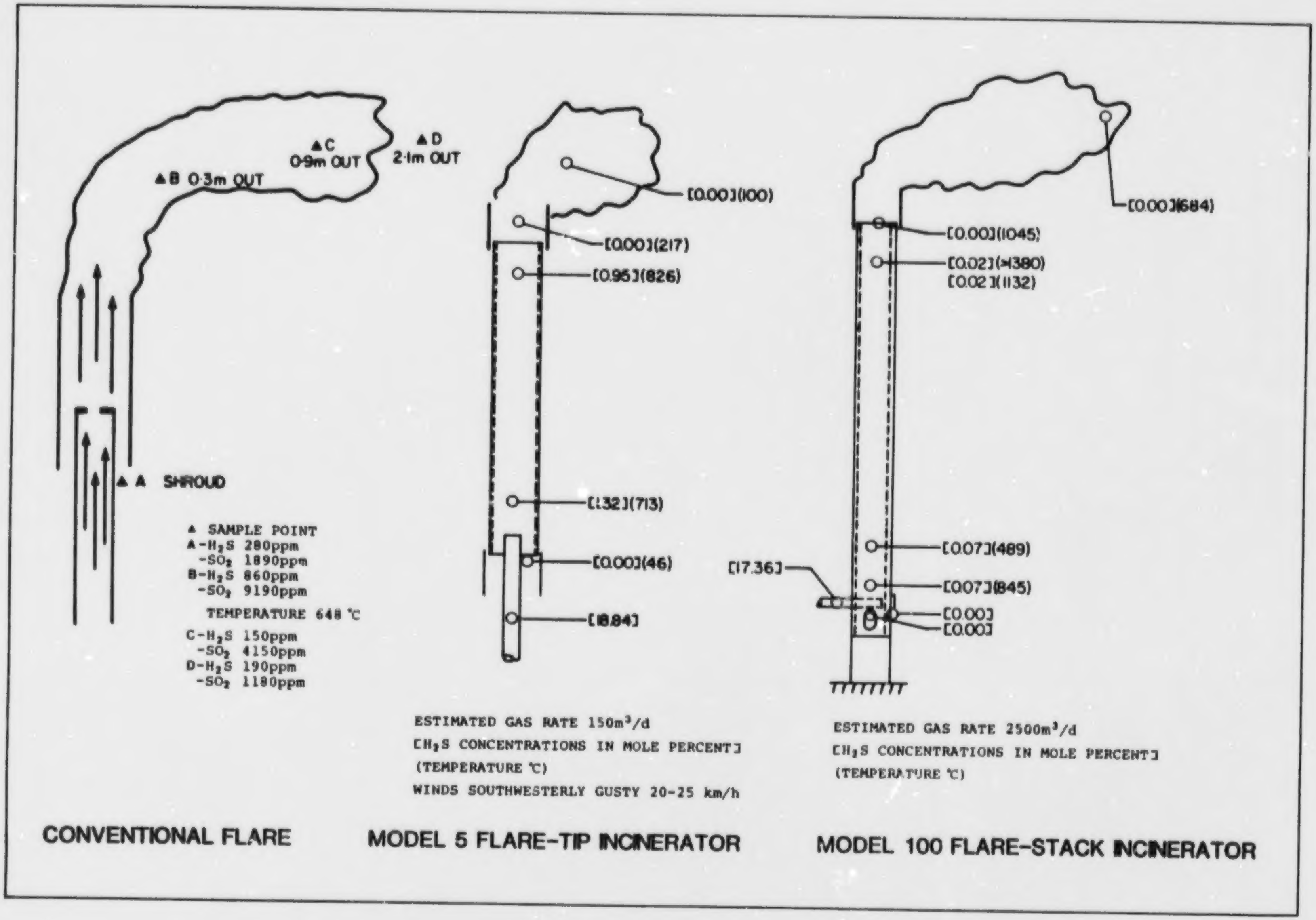

▲D
2·1m OUT
▲C
0·9m OUT
▲B 0·3m OUT
▲A SHROUD
▲ SAMPLE POINT
A-H₂S 280ppm
-SO₂ 1890ppm
B-H₂S 860ppm
-SO₂ 9190ppm
TEMPERATURE 648 °C
C-H₂S 150ppm
-SO₂ 4150ppm
D-H₂S 190ppm
-SO₂ 1180ppm
[0.00](100)
[0.00](217)
[0.95](826)
[1.32](713)
[0.00](46)
[18.84]
ESTIMATED GAS RATE 150m³/d
[H₂S CONCENTRATIONS IN MOLE PERCENT]
(TEMPERATURE °C)
WINDS SOUTHWESTERLY GUSTY 20-25 km/h
[0.00](684)
[0.00](1045)
[0.02](>1380)
[0.02](1132)
[0.07](489)
[0.07](845)
[17.36]
[0.00]
[0.00]
ESTIMATED GAS RATE 2500m³/d
[H₂S CONCENTRATIONS IN MOLE PERCENT]
(TEMPERATURE °C)
CONVENTIONAL FLARE
MODEL 5 FLARE-TIP INCINERATOR
MODEL 100 FLARE-STACK INCINERATOR

FIGURE 13: TYPICAL TEST RESULTS OF CONVENTIONAL FLARING AND INCINERATION

SOURCE: (21)

sulphuric acid by bacterial action. The rate of sulphuric acid formation depends on the particle size of the sulphur dust, the temperature, moisture, and the soil type. Under favourable conditions sufficient acid may be generated to acidify most type of soils to such an extent that the indigenous vegetation may be affected. At the soil pH below 5.5, toxic trace elements, such as aluminum and manganese, could be made available to the growing plants(22).

Some concerns regarding low soil pH and unnatural sulphur/selenium balance in the soil around certain sour gas processing plants with sulphur storage facilities have been reported(23). Apparently, the deposition of sulphur dust is regarded as one of the causes for the above phenomena. Fugitive dust emission from sulphur storage and handling area can be kept to a minimum by using proper enclosures and hooded loading-unloading facilities.

During the past few years the existence of sulphur dust deposition problems adjacent to sulphur storage areas has been recognized in the province of Alberta, and several investigations have been undertaken(22,24).

6.1.3. Acid Gas Disposal or Treatment

The acid gas consists mostly of CO_2 and sulphur compounds (mainly H_2S with traces of COS, CS_2 and RSH) separated during the processing of raw sour gas. While the major source of acid gas is the sour gas desulphurization system, other minor sources include hydrocarbon recovery and sour effluent treatment units. Analyses of inlet gas and acid gas from three processing plants are shown in Table 9.

TABLE 9

ANALYSIS OF INLET GAS AND ACID GAS

(mole %)

Constituent	Fort Nelson Plant		Taylor Plant		Pine River Plant	
	Inlet Gas	Acid Gas	Inlet Gas	Acid Gas	Inlet Gas	Acid Gas
CO_2	12.4	85.3	1.8	54.0	9.9	44.0
H_2S	1.8	13.6	1.5	45.0	14.2	54.0
C_1	85.5	0.6	85.4	0.6	75.4	1.0
C_2	0.3		6.6		0.2	
C_3			2.5			
C_4^+			2.2			
N_2		0.5*		0.4*	0.3	1.0*

*Also contains traces of COS and CS_2

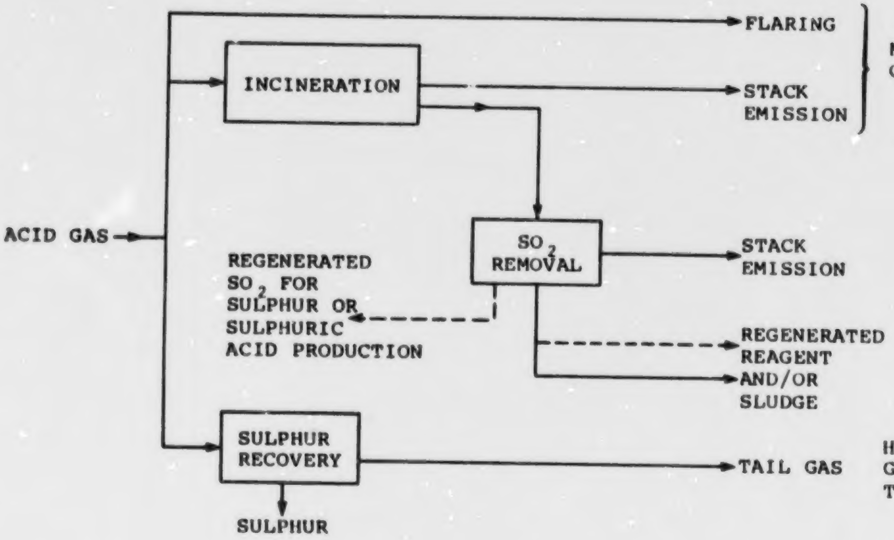
(Source: Westcoast Transmission Company Ltd.)

From environmental standpoint the method of acid gas disposal or treatment is considered to be the most significant aspect of the gas processing industry. There are several options available for this purpose as shown in Figure 14. The choice of a particular system would depend mainly on the sulphur content of the acid gas and the pollution control regulations governing the limits on SO_2 emission and sulphur recovery. It may range from flaring through a stack at a small installation to recovery of maximum amount of sulphur from the acid gas in a central processing plant.

In smaller installations or at well-heads any acid gas generated may be disposed of by flaring provided the SO_2 emitted is within the regulatory limits. As discussed in a previous section, the acid gas could be incinerated instead of flaring for better combustion and lesser risk of emitting unconverted H_2S into the atmosphere.

In cases where the quantity and quality of the acid gas precludes any sulphur recovery and its incineration creates SO_2 emission problems, it may necessitate removal of SO_2 from the incinerated gas prior to its discharge. The degree of SO_2 removal and the choice of the flue gas desulphurization (FGD) method would be site-specific and governed by various factors.

There are several FGD methods available commercially, and a number of processes are under various stages of development. The information currently available on FGD is quite large and could be the subject of a separate report. For the present purpose key features of four commercially available processes are listed in Table 10. In British Columbia, no FGD process


ACID GAS
FLARING
INCINERATION
STACK EMISSION
SO_2 REMOVAL
STACK EMISSION
REGENERATED SO_2 FOR SULPHUR OR SULPHURIC ACID PRODUCTION
REGENERATED REAGENT AND/OR SLUDGE
SULPHUR RECOVERY
TAIL GAS
SULPHUR

MAY BE UNACCEPTABLE DUE TO HIGH SO_2 CONCENTRATION IN THE EMITTED GAS

H_2S CONCENTRATION IN THE TAIL GAS MAY BE HIGH. TAIL GAS TREATMENT MAY BE NECESSARY

FIGURE 14: OPTIONS FOR ACID GAS TREATMENT AND DISPOSAL

TABLE 10 - KEY FEATURES OF FOUR FLUE GAS DESULPHURIZATION PROCESSES

Features	Lime/Limestone	Dual Alkali	Chiyoda Thoroughbred 101	Wellman-Lord
Principle	Absorption of SO_2 in a lime or limestone slurry (6-12% solids); throw-away process	Absorption of SO_2 in a sodium hydroxide, sodium sulphite, sodium sulphate and sodium carbonate solution; regenerative process	Absorption of SO_2 in 2% (w/w) sulphuric acid solution; regenerative process	Absorption of SO_2 in a concentrated sodium sulphite-bisulphite solution; regenerative process
Feed Stream Requirements	-	Excessive particulates must be removed; O_2 concentration must be less than 7%	Particulates removal necessary	Particulates removal necessary
Efficiency	Depends on scrubber type, gas pressure drop and SO_2 concentration in feed gas. Generally 70 to 90% SO_2 removal is achieved, but over 90% can be achieved	Capable of removing over 90% SO_2	95% SO_2 removal can be achieved	90% + SO_2 removal can be achieved
Product	Calcium sulphite and calcium sulphate	Calcium sulphite and calcium sulphate as products of absorbent regeneration	Gypsum with 20-50% moisture	Concentrated SO_2 stream for sulphuric acid or elemental sulphur production
Advantages/ Disadvantages	Low capital and O & M costs, commercially proven process. Produces large amount of sludge, sludge fixation may be necessary, on-line reliability may be low	Low capital and O & M costs, commercially proven process. Produces some sludge, may present equipment corrosion problem	Produces saleable products, commercially demonstrated process. Utility requirements high, special metals must be used in equipment construction because of sulphuric acid use	Lower potential for fouling/scaling than Ca-based processes, commercially demonstrated. Utility requirements high, requires separate process for SO_2 disposal

is used to remove SO_2 from incinerated acid gas, as in most cases the acid gas is treated for sulphur recovery.

6.1.3.1. Sulphur Recovery

A number of processes are available commercially for recovery of sulphur from the acid gas. These can be classified into four major categories:

1. Wet-bed Absorption (air oxidation)
2. Wet-bed Absorption (direct conversion)
3. Dry-bed Absorption
4. Dry-bed Catalytic Conversion

Examples of some of the processes under categories 1 and 2 are listed as Direct Conversion Process and that of category 3 as Dry-bed Process in Table 5. The Claus (modified) and Direct Oxidation (Pan American Petroleum Corp.) processes fall under category 4.

The choice of a particular system would depend on feed gas characteristics as well as other factors. In the gas industry, however, the major governing factors seem to be the H_2S content in the acid gas and the extent of sulphur removal necessary to meet the environmental regulations. The general features of three processes are shown in Table 11, and it is obvious that to treat an acid gas containing more than 2% H_2S the Claus process is the most suitable. The sulphur recovery efficiency of the Claus process increases with increasing H_2S content in the feed gas. Acid gases with low H_2S content (2-15% mole) can be treated by Direct Oxidation process with a sulphur recovery efficiency between 80 and 85% only(14,27). The Claus process is used for sulphur recovery at the gas processing plants in British Columbia.

TABLE 11 - GENERAL FEATURES OF SULPHUR RECOVERY PROCESSES

Process	Principle of Process	Feed Characteristics	Removal Efficiency, %						By-product	Effect of CO_2	Reference
			H_2S	COS/CS_2	RSH	HCN	NH_3	HC			
Stretford	Liquid phase oxidation of H_2S to elemental sulphur. Alkaline solution of metavanadate and anthraquinone disulphonic acid salts are used as oxidation medium	Presently applied to 1% or less total sulphur in the feed	99.9	0	0	100	0	0	Elemental sulphur	High CO_2 concentration in the feed decreases absorption efficiency by lowering the alkalinity of the solution	20, 29-31
Claus	Catalytic oxidation of H_2S and SO_2 to elemental sulphur	Widely applied for feed gas containing 2% or more H_2S using different modes of operation	90-95	90	95	Partial	Partial	90	Elemental sulphur	High CO_2, HC and NH_3 content can adversely affect sulphur removal capacity, and can cause catalyst fouling	11, 25-28
Giammarco-Vetrocoke (G-V)	Liquid phase oxidation of H_2S to elemental sulphur. Alkaline potassium carbonate and arsenate/arsenite solution is used as oxidation medium	Maximum of 1.5% H_2S in feed gas	99.99	Partial	Partial	?	0	0	Elemental sulphur which may need	Little or no effect	17, 18. 32-33

Claus Process

The process was originally developed about 1880 by Claus when he produced elemental sulphur by catalytic oxidation of H_2S according to the following overall reaction.

$$2H_2S + O_2 \rightleftharpoons 2S + 2H_2O + \text{heat} \qquad (1)$$

Because of the exothermicity and reversibility of the above reaction, lower-temperature operations favour conversion to sulphur. However, at too low a temperature, catalyst poisoning occurs by deposition of elemental sulphur. Formerly, the major operational problem experienced was the temperature control in the catalytic converter. To surmount the problem of high temperature, over-designed catalytic converters were used; even so the sulphur recovery yield was about 60% and considered to be poor(11).

Modified Claus Process

Subsequent work at the I.G. Farben Industrie in Germany led to the development of a modified Claus process in about 1937. Much higher efficiency of conversion and recovery of sulphur was obtained by exploiting the fact that the overall reaction (1) proceeds in two distinct stages:

$$2H_2S + 3O_2 \rightleftharpoons 2SO_2 + 2H_2O + \text{heat} \qquad (2)$$

$$2H_2S + SO_2 \rightleftharpoons 3S + 2H_2O + \text{heat} \qquad (3)$$

where $3S = xS_2 + yS_6 + zS_8$ and s, y, and z represent the number of moles of elemental sulphur in the forms of S_2, S_6, and S_8 respectively.

In the modified Claus process, a portion of the H_2S is oxidized to SO_2 (Reaction 2) in a reaction furnace. The products of reaction are cooled to recover the heat of reaction as well as to condense out any elemental sulphur formed in the reactor by Reaction (1). After adjusting its temperature to the level necessary for optimum conversion to sulphur, the uncondensed gases are reacted with the remaining H_2S in a catalytic reactor to produce elemental sulphur by Reaction (3). Due to the exothermicity of the reaction, the sulphur conversion efficiency of a single-stage unit is usually limited. Accordingly, additional stages are frequently used in series, with interstage cooling of the gas and condenstion of sulphur, to achieve the required extent of sulphur removal.

For a multi-component feed, containing CO_2 and CH_4, several side reactions occur in the reaction furnace to produce carbonyl sulphide (COS) and carbon disulphide (CS_2) as follows:

$$CO_2 + H_2S \rightleftharpoons COS + H_2O \qquad (4)$$
$$COS + H_2S \rightleftharpoons CS_2 + H_2O \qquad (5)$$
$$CH_4 + 4S \rightleftharpoons CS_2 + 2\ H_2S \qquad (6)$$

Such productions of organic sulphur compounds represent a loss of about 0.25 to 1% of the feed sulphur depending on the particular design of the process(11,34). More detailed discussion on the Claus process chemistry and data on the kinetics and yield of various reactions can be found in a paper by Gamson and Elkins(35).

Process Variations

Since the introduction of the modified Claus process, and especially after 1950, several variations of the process have been developed. Primarily based on the H_2S content of the feed acid gas, the following basic schemes are available today:

1. Straight-through;
2. Split-stream;
3. Sulphur recycle; and
4. Direct oxidation.

Although variations of each of the above schemes are practised to accommodate different feed gas characteristics and sulphur recovery level desired, the basic configurations are shown in Figure 15. For an acid gas containing a high H_2S concentration (normally 50%+), the straight-through scheme is reportedly most suitable. In this scheme, the entire feed gas is sent to the combustion furnace with sufficient air to oxidize about one-third of the H_2S to SO_2. Approximately 50 to 60% conversion of the H_2S to sulphur occurs at this stage when the reaction temperature approaches nearly 1 090°C. The hot gases from the furnace are passed through a waste heat boiler and a sulphur condenser to separate elemental sulphur. The condenser outlet gas at about 145°C is reheated by hot, by-pass gas to about 250°-320°C before being passed through the first catalytic converter to produce sulphur via Reaction (3). The sulphur produced is separated from the gas in a condenser, and the residue gas is reheated and passed through a second catalytic

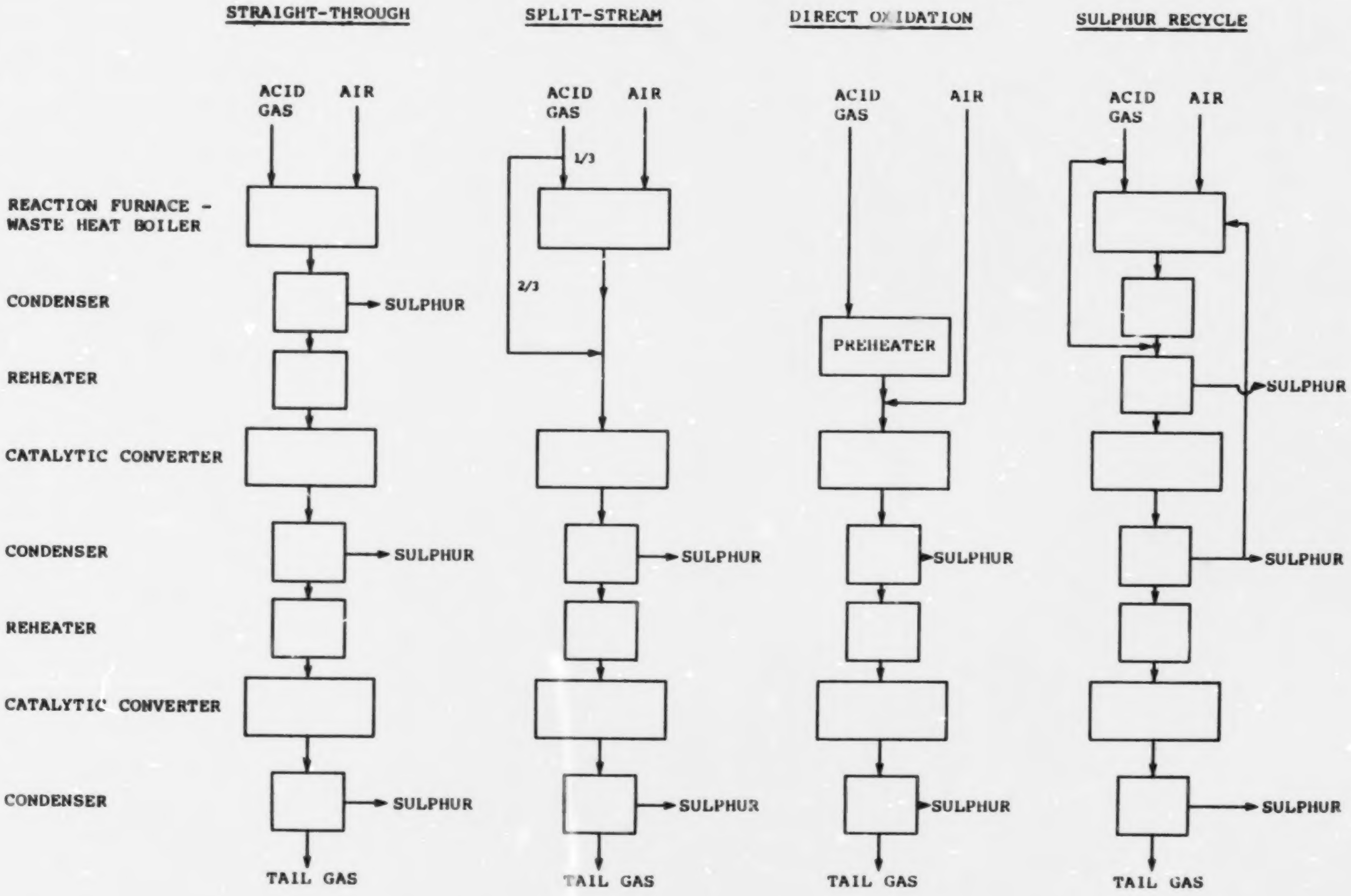

STRAIGHT-THROUGH
SPLIT-STREAM
DIRECT OXIDATION
SULPHUR RECYCLE
ACID GAS
AIR
1/3
2/3
PREHEATER
REACTION FURNACE - WASTE HEAT BOILER
CONDENSER
REHEATER
CATALYTIC CONVERTER
CONDENSER
REHEATER
CATALYTIC CONVERTER
CONDENSER
SULPHUR
TAIL GAS

FIGURE 15: BASIC CONFIGURATIONS OF MODIFIED CLAUS PROCESS

converter to facilitate further reaction between SO_2 and H_2S to produce additional sulphur. After condensation and removal of the sulphur, the tail gas from the second-stage condenser may be further processed in one or more catalytic converters or disposed of by one of the methods discussed later.

The split-stream scheme is best suited for acid gas streams with H_2S content in the range of 15 to 60%. In this configuration one-third of the acid gas stream is fed to the furnace with sufficient air for complete combustion of H_2S to SO_2. After being cooled in a waste heat boiler, the product gas is mixed with the remaining two-thirds and sent to the first catalytic converter for the production of sulphur. The rest of the scheme is the same as described for the straight-through scheme.

For the treatment of acid gases containing medium to high levels of H_2S, the straight-through configuration is generally superior to the split-stream version as shown in Table 12. However, the final choice would be governed by other factors such as feed gas quality, the degree of sulphur removal necessary, etc.

For lean acid gases with 2 to 15% H_2S, sulphur recycle or direct oxidation configuration can be applied. In the former scheme, a portion of the product sulphur is recycled to the furnace to generate the desired quantity of SO_2 for reaction with the H_2S present in the acid gas. The SO_2 generated in the furnace is mixed with the acid gas stream before the catalytic converter. The direct oxidation scheme is based on Reaction (1), but is practised nowadays with better process control. In this scheme air is mixed with the preheated gas stream, and passed directly through the catalyst bed in the oxidation

TABLE 12: COMPARISON OF MODIFIED CLAUS PROCESS SCHEMES

	Straight-through	Split-stream
Investment:		
Treating plant	Lower	Higher
Sulphur plant	Lower	Higher
Operating Cost:		
Fuel, maintenance, etc.	Lower	Higher
Sulphur Recovery Level	Higher	Lower
Downtime	Lower	Higher
Control of plant stream composition	Easier	More difficult

Source: (34)

reactor. The separation of sulphur from the outlet gas is usually practised in manners similar to those in other two schemes.

In British Columbia, straight-through method is used at Taylor and Pine River plants, and sulphur recycle configuration is used at Fort Nelson because of low H_2S content in the acid gas produced.

Emissions from Claus Plant and Controls

The outlet gaseous stream from the final coalescer in a Claus sulphur recovery plant, commonly termed as the tail gas, contains several pollutants including H_2S, reduced S-compounds such as COS and CS_2, SO_2, CO, and volatile organic compounds. When no additional control is used, the tail gas is usually incinerated and discharged through a stack. The stack gas emission consists mostly of SO_2 as the major pollutant along with some small amount of CO. However, to comply with the emission limits for SO_2, it is often necessary to treat the tail gas before discharge. The options for the disposal of Claus plant tail gas are discussed in a later section.

From the preceding discussion it becomes obvious that the key to the presence of a minimum amount of sulphur compounds in the tail gas is the maximum sulphur recovery efficiency of the Claus plant. There are several factors which affect this efficiency, but the major ones are as follows:

- the process chemistry
- feed gas composition

- control of the proper stoichiometric ratio of $H_2S:SO_2$
- number of catalytic conversion stages
- control of operating temperatures
- catalyst maintenance and plant capacity

Because of the reversibility of Reaction (3), complete conversion of H_2S to S and water can never be achieved, as these products are always present in the reactor in sufficient amounts to initiate the reverse reaction.

Moreover, there are some practical operating limitations which preclude achieving theoretical conversion efficiency in the Claus plant. The loss of sulphur via various routes is the major factor in attaining theoretical recovery. Typical examples of sulphur losses from a straight-through two-stage plant is shown in Table 13. Some of these losses can be minimized by measures taken to control the adverse effects of other factors discussed below.

The composition of the acid gas feed has direct bearing on the sulphur recovery efficiency. As shown in Table 14, for a given number of catalytic converters, sulphur recovery increases as the H_2S content of the feed gas increases(36). However, the presence of hydrocarbons, high CO_2, water vapour, ammonia, and carry-over gas treating solution affects the sulphur recovery efficiency mostly by reducing catalyst life(26,34).

Organic compounds, in forms of hydrocarbon gases and treating solution carry-over, entering the reaction furnace with the feed require additional air for combustion, and produce increased amounts of water vapour,

TABLE 13: TYPICAL LOSS OF SULPHUR FROM A STRAIGHT-THROUGH TWO-STAGE CLAUS PLANT

Loss/Recovery	Per cent of inlet sulphur
Loss	
As $H_2S + SO_2$	2.70
As $COS + CS_2$	0.35
As sulphur vapour	0.25
As entrained sulphur	0.50
Recovery	96.20

Source: (34)

TABLE 14: TYPICAL RANGE OF SULPHUR RECOVERY FOR VARIOUS H_2S CONTENTS OF THE CLAUS PLANT FEED GAS

H_2S content of feed gas % (dry basis)	Calculated Sulphur Recovery, %		
	Two Converters	Three Converters	Four Converters
20	92.7	93.8	95.0
30	93.1	94.4	95.7
40	93.5	94.8	96.1
50	93.9	95.3	96.5
60	94.4	95.7	96.7
70	94.7	96.1	96.8
80	95.0	96.4	97.0
90	95.3	96.6	97.1

Estimates based on allowances for normal organics carry-over and entrainment

Source: (36)

CO_2 and N_2 in the furnace. These constituents act as diluents and shift the reaction equilibrium adversely affecting sulphur conversion. Carbon dioxide and methane take part in the unwanted side reactions (Reactions (4) - (6)) forming COS and CS_2 which are difficult to convert to elemental sulphur. In addition, high CO_2 concentration reduces the catalyst life, and organics of higher molecular weight encourage the formation of soot which lowers the catalyst efficiency.

The presence of ammonia, even in the range of 500-1000 ppm_v, affects the Claus plant operation by catalyst fouling, plugging of process lines and condenser tubes (by ammonium sulphate and ammonia hydrosulphides), and by production of dark-coloured sulphur(22,26).

It is obvious that proper front end design to control the Claus plant feed gas composition is essential for increased sulphur recovery and reduced sulphurous emissions. The concentrations of hydrocarbons and water vapour in the acid gas is normally controlled by installing an amine flash tank. Additional control of these constituents can be achieved by reducing the amine circulation rate in the absorber, and by controlling temperatures of both the inlet gas and amine. The water vapour content can also be reduced by controlling the temperature of the amine regenerator reflux accumulator(34).

To control excessive CO_2 in the acid gas the choice of desulphurization solvent is critical. Selective absorption of H_2S by the solvent to generate an acid gas with low CO_2/H_2S ratio is desirable. There are commercially developed solvents, such as MDEA, available which absorbs virtually all of the H_2S but only a portion of the CO_2 present in the sour gas(37).

Ammonia is usually removed upstream of the Claus plant. Its concentration in the acid gas stream can be reduced to an acceptable level (less than 500 ppm_v) by ensuring adequate combustion in the reaction furnace. Alternatively, modifications to the operating conditions could be carried out to prevent interference by ammonia(26,34).

According to Reaction (3), it is essential that the initial ratio of $H_2S:SO_2$ in the catalytic reactor is maintained at the stoichiometric ratio of 2:1 for a maximum conversion to sulphur. To achieve this ratio, an exact amount of air must be supplied to the acid gas reaction furnace. The effect of initial $H_2S:SO_2$ ratio on the conversion to sulphur is shown in Figure 16. The ratio of $H_2S:SO_2$ is controlled by the quantity of combustion air supplied to the reaction furnace. Because of practical reasons the optimum ratio of $2.0(H_2S/SO_2)$ is difficult to maintain, and the usual practice is to supply the combustion air slightly in excess of the stoichiometric amount. As shown in Table 15 the loss in sulphur recovery is less when an excess air is present. The table also reveals that for efficient conversion the air supply should be maintained within + 2% in the case of two-converter plant and within + 1% for a three-converter plant.

Presently due to improvements in continuous gas analysers and telemetry system, it is possible to make fine adjustments to the air supply rate for any changes in the acid gas quantity and quality so that $H_2S:SO_2$ ratio can be maintained continuously very close to 2.0(26,39).

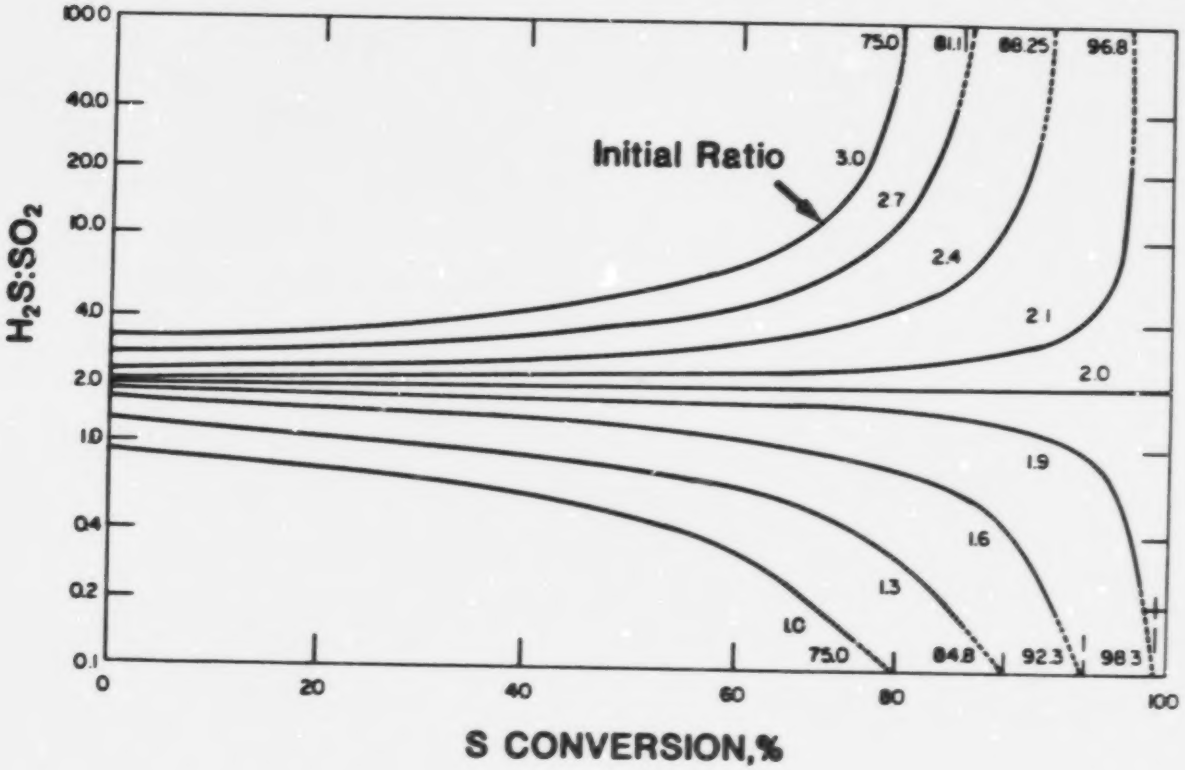


FIGURE 16: VARIATION OF $H_2S:SO_2$ RATIO WITH CONVERSION TO SULPHUR AND MAXIMUM THEORETICAL CONVERSION AT DIFFERENT RATIOS

SOURCE: (38)

TABLE 15: EFFECT OF QUANTITY OF AIR SUPPLY TO THE REACTION FURNACE ON ($H_2S + SO_2$) LOSS FROM CLAUS REACTOR

Air Supply % Stoichiometric Air	($H_2S + SO_2$) Loss, As % H_2S in Claus Inlet Gas	
	Two Converters	Three Converters
97	3.60	3.10
98	3.12	2.14
99	2.70	1.32
100	2.53	1.05
101	2.56	1.20
102	2.79	1.54
103	3.20	2.10

Other sulphur losses are assumed to be constant.

Source: (34)

For a given H_2S content in the Claus plant feed gas, the sulphur recovery efficiency increases with the number of catalytic converters, as shown in Table 14 and Figure 17. However, the presence of water vapour and the progressive reduction in the H_2S content of the gas make each additional converter less efficient than the preceding one.

As the Claus reactions are exothermic, lower operating temperatures in the reactors favour sulphur conversion efficiency. However, this temperature is limited in practice at the range of 200°-260°C to avoid condensation of sulphur on the catalysts. The sulphur condensers should also be operated at the lowest practical temperature (around 127°C) which would be governed by the desired viscosity of the liquid sulphur, avoidance of its solidification, and condensation of water. By maintaining a lower temperature in the first catalytic converter, about 350°C, a major portion (40-70%) of the COS formed in the waste heat boiler tends to reconvert to H_2S according to Reaction (4).

The catalytic converter design and the size and shape of the catalyst should be such that maximum contact between the fluid and the bed is maintained. Any kind of catalyst fouling should be avoided to maintain its conversion efficiency. It has also been reported that the sulphur recovery efficiency is somewhat reduced by operating the plant below capacity(41).

Because of the factors mentioned above, even a properly designed and operated Claus plant will achieve only 93-97% sulphur recovery efficiency(42). The higher figure could be attained with a fourth catalytic converter and under the best of operating conditions. The

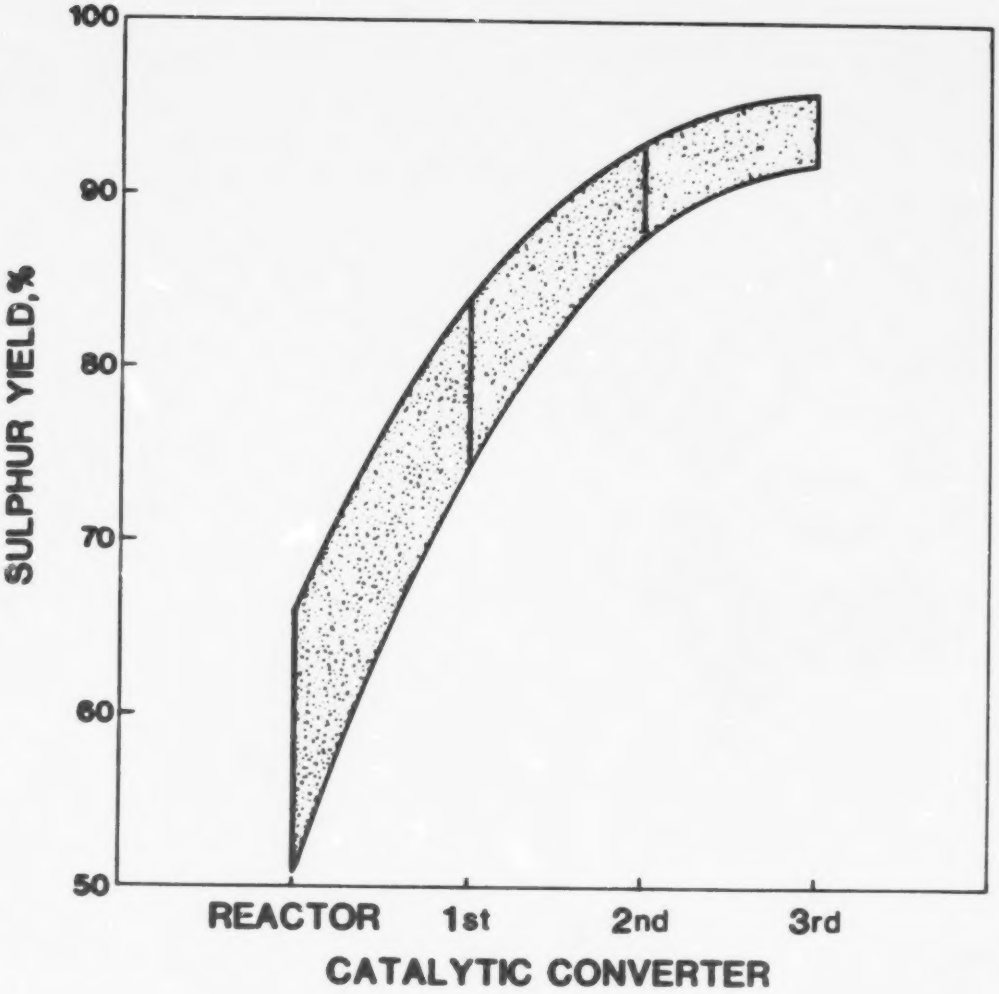


FIGURE 17: EFFECT OF THE NUMBER OF CATALYTIC CONVERTER ON SULPHUR RECOVERY (ASSUMING 1% (MOLE/MOLE) OF HYDROCARBONS IN THE FEED AND CATALYST TEMPERATURES OF 204°-260°C)

SOURCE: (40)

tail gas from a Claus plant, containing varying proportions of H_2S, COS, CS_2, S-Vapour, and SO_2, is conventionally incinerated to convert all forms of sulphur into SO_2 and discharged through a stack. Typical SO_2 concentration from such a plant is in the range of 10 000-20 000 ppm_v, as shown in Figure 18, depending on the H_2S content of the acid gas and sulphur recovery efficiency(36). To minimize the effects of such a high concentration of SO_2, the usual practice has been to design the incinerator stack height for sufficient dispersion of the gas and maintain the ground level SO_2 concentration in the plant vicinity within the air quality regulatory limits. However, such practice results in a significant quantity of sulphur emission from medium and large plants. For example, after a recovery of 95% of the inlet sulphur from a plant with a feed rate of 1 000 t/d, there would be an emission of 50 t/d of sulphur or 100 t/d of SO_2 from the incinerator stack.

Considering this aspect as well as the potential long-term effects of SO_2 emission, the regulatory agencies are stipulating the required sulphur recovery based on the plant size and sour gas quality to effectively control SO_2 emission at the source. To comply with such requirements, the overall sulphur conversion and recovery efficiency of a Claus plant must be in the range of 99%+, which would necessitate some form of tail gas treatment downstream of the sulphur recovery plant.

6.1.3.2. Tail Gas Treatment

An analysis of a typical Claus plant tail gas is shown in Table 16, and as can be seen it produces a high concentration of SO_2 in the incinerated gas. With the exception of small plants, where the total sulphur

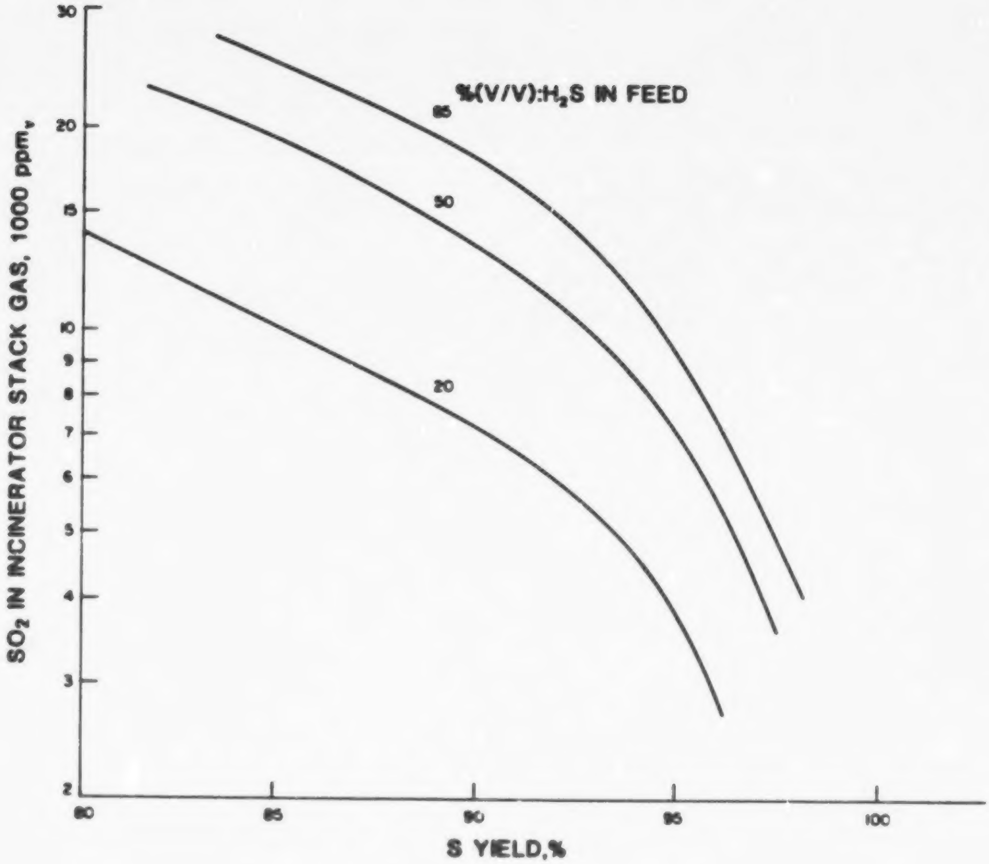


FIGURE 18: SULPHUR DIOXIDE CONCENTRATION IN INCINERATOR STACK GAS AS A FUNCTION OF SULPHUR YIELD AND HYDROGEN SULPHIDE CONTENT IN THE FEED TO A CLAUS PLANT (ASSUMING SUFFICIENT FUEL GAS ADDITION AND 25% EXCESS AIR TO MAINTAIN A MINUMUM TEMPERATURE OF 538°C AT THE TOP OF THE STACK)

SOURCE: (36)

TABLE 16: TYPICAL CLAUS PLANT TAIL GAS COMPOSITION

Constituent	mode % (wet basis)
H_2S	1.213
SO_2	0.192
COS	0.254
CS_2	0.098
H_2	0.006
CO	0.318
CO_2	8.371
N_2	60.020
Ar	1.716
CH_4	0.152
S_{vapour}	0.006
S_{vapour}	0.020
Moisture	27.634
Total	100.00

After a 3-stage Claus reactor and 96% sulphur recovery.

Total SO_2 to be generated on incineration is approximately 14 000 ppm_v in the stack gas, or about 80 t/d from a sulphur recovery plant with 1 000 t/d of sulphur in the feed gas.

emission is not large enough to install an economic treatment system, some form of tail gas treatment is practised nowadays to reduce its sulphur content to an acceptable level.

Additional sulphur removal from the tail gas is possible by three general schemes as shown in Figure 19. Approximately seventy processes under various stages of development and with potential for application in the tail gas treatment have been reported in the literature. Brief descriptions of major features of some commercially available processes are given in Table 17.

It should be noted that there are several practical limitations in the treatment of a gaseous stream with low H_2S concentration. Any add-on system will most likely produce some kind of waste streams and their disposal problems, and need additional energy for heating and cooling. It may also require additional measures for proper operation of the upstream Claus plant. As a result the tail gas treatment methods become expensive, and the overall cost of any incremental recovery of sulphur beyond 99% is extremely high. The sulphur recovery capabilities of some of these processes in conjunction with two-stage Claus plants are summarized in Table 18.

In British Columbia only the three major plants have sulphur recovery facilities. As indicated in Table 19 only two of the major plants also practise tail gas treatment, while at the other the Claus plant tail gas is incinerated. The treatment of the tail gas at the Fort Nelson and Pine River plants is carried out by the same basic principle of dry medium extension of the Claus reaction under sub-dew point temperature. However, the actual process used at the former plant is the Sulfreen

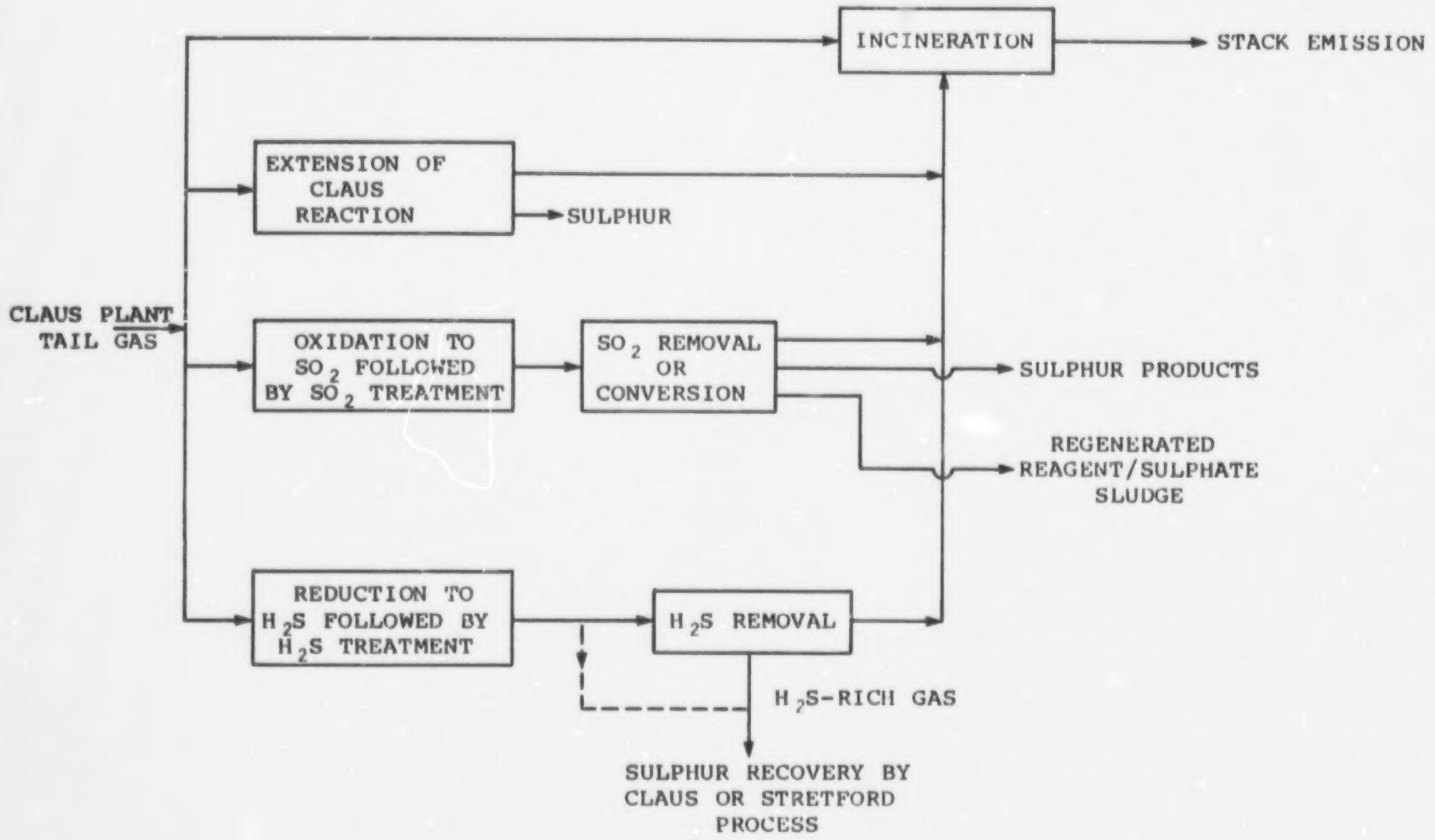

CLAUS PLANT
TAIL GAS
INCINERATION
STACK EMISSION
EXTENSION OF CLAUS REACTION
SULPHUR
OXIDATION TO SO_2 FOLLOWED BY SO_2 TREATMENT
SO_2 REMOVAL OR CONVERSION
SULPHUR PRODUCTS
REGENERATED REAGENT/SULPHATE SLUDGE
REDUCTION TO H_2S FOLLOWED BY H_2S TREATMENT
H_2S REMOVAL
H_2S-RICH GAS
SULPHUR RECOVERY BY CLAUS OR STRETFORD PROCESS

FIGURE 19: CLAUS PLANT TAIL GAS TREATMENT OPTIONS

TABLE 17: MAJOR FEATURES OF SOME CLAUS PLANT TAIL GAS TREATMENT SYSTEMS

Basic Principle	Process Medium	Process Name	Process Description	Efficiency	Product	Reference
Extension of Claus reaction for additional sulphur recovery	Dry	Amoco CBA (Cold Bed Adsorption)	Tail gas is passed through an adsorption reactor containing Claus catalyst at lower temperature (125°-150°C) to shift equilibrium reactions towards sulphur formation. Sulphur adsorbed by the catalyst is regenerated by hot gas while reactivating the catalyst. COS and CS_2 are not reduced.	With 2-stage Claus and 1-stage CBA, overall recovery of sulphur up to 99%. Final incinerated gas will contain 1 500-2 500 ppm_v of SO_2.	Elemental sulphur	41, 43
		Sulfreen	Similar to above with the primary difference in the catalyst regeneration method. While the Sulfreen system uses a closed loop method, the CBA is operated with an open cycle. COS and CS_2 are not reduced.	Similar efficiency as for the CBA.	Elemental sulphur	44-46
	Wet	IFP (Institut Francais du Petrole)	Liquid phase catalytic conversion of H_2S to S at a low temperature (121°-132°C). H_2S:SO_2 ratio is maintained at about 2.0. COS and CS_2 are not converted.	Capable of reducing SO_2 content in the incinerated stack gas to as low as 1 000ppm_v. With 2-stage Claus plant (95% recovery) and IFP, overall sulphur recovery up to 98.4%.	Elemental sulphur	48-50
Oxidation to SO_2 followed by SO_2 treatment	Dry	Shell SFGD (Shell Flue Gas Desulphurization)	Thermal oxidation to SO_2 and adsorption by copper oxide-on-alumina at about 400°C. SO_2 regenerated by a hot reducing gas (H_2 or H_2-CO mixture).	90% of SO_2 is removed. Treated gas may contain 1 500-3 000 ppm_v of SO_2.	Concentrated SO_2 stream for sulphuric acid or other by-products	51-52
		SNPA Topsoe (Societe Nationale des Petroles d'Aquitaine Haldor Topsoe)	Catalytic oxidation to SO_2 and then conversion of SO_2 to sulphuric acid.	95% of SO_2 is removed.	Sulphuric acid.	36,52-53
	Wet*					
Reduction to H_2S followed by H_2S treatment	Wet	SCOT (Shell Claus Off-gas Treating)	Reduction of sulphur compounds to H_2S by H_2 over a Co-Mo catalyst at 300°C. H_2S is first absorbed by an alkanolamine solvent, and then the desorbed H_2S-rich stream is sent to the Claus plant. High CO_2 level affects both catalytic conversion and H_2S absorption by the solvent.	Can remove up to 97% of sulphurous compounds and produce a vent gas from the H_2S absorber containing 250-500 ppm_v.	Concentrated H_2S stream for Claus plant.	54-57
		Beavon	Catalytic reduction of all sulphur compounds to H_2S by H_2, followed by Stretford process to recover elemental sulphur. Recently a modified version of the process has been introduced to replace Stretford process by a catalytic oxidation process known as Selectox to convert H_2S to S.	An overall sulphur recovery up to 99.9% can be achieved with a final tail gas containing less than 250 ppm_v of SO_2.	Elemental sulphur	58-60
		Trencor-M	The Claus tail gas is hydrogenated at about 300°C over a cobalt-molybdenum catalyst to reduce all sulphur compounds to H_2S. A proprietary solution of MDEA is used to absorb H_2S from the reduced gaseous stream. The regenerated H_2S is recycled to the Claus plant, and the regenerator vent gas is incinerated.	The incinerator stack gas reportedly contains SO_2 in the range of 100-250 ppm_v.	Elemental sulphur	52, 61

* The processes under this category are listed in Table 9.

TABLE 18: OVERALL SULPHUR RECOVERY CAPABILITY OF SOME TAIL GAS CLEAN-UP SYSTEMS IN CONJUNCTION WITH CLAUS PLANT

System	Sulphur Recovery Capability, %	Reference
2-Converter Claus (96% recovery) with 1-converter CBA	99	43,52,62
2-Converter Claus (96% recovery) with 2-converter CBA	99.5	43,52,62
2-Converter Claus (95% recovery) with 1-converter Sulfreen	98.6	44,46,52,62
2-Converter Claus (95% recovery) with 2-converter Sulfreen	99	44,46,52,62
2-Converter Claus (95% recovery) with IFP	98.1-99.4	49,52
2-Converter Claus (94% recovery) with SCOT	99.85	42,54
2-Converter Claus (93-96% recovery) with Beavon	99.99	42,58-60
2-Converter Claus (94-96% recovery) with Wellman-Lord	99.9	42,62-63

TABLE 19

EXISTING SULPHUR RECOVERY PLANTS (MODIFIED CLAUS) IN BRITISH COLUMBIA

Plant	Location	Process Mode	Plant Configuration	Tail Gas Treatment and Disposal	Overall Sulphur Recovery, % (Designed)
Petro-Canada Exploration Inc. and Westcoast Transmission Co. Ltd.	Taylor	Straight-Through	3-stage Claus (Two trains of 2-stage plus a common stage)	Incineration	95
Westcoast Transmission Co. Ltd.	Fort Nelson	Sulphur Recycle	Two trains of 1-stage Claus	One common Sulfreen unit followed by incineration	95
Westcoast Transmission Co. Ltd.	Pine River (Chetwynd)	Straight-Through	Two trains of 2-stage Claus	Two trains of 2-stage CBA unit followed by incineration	99

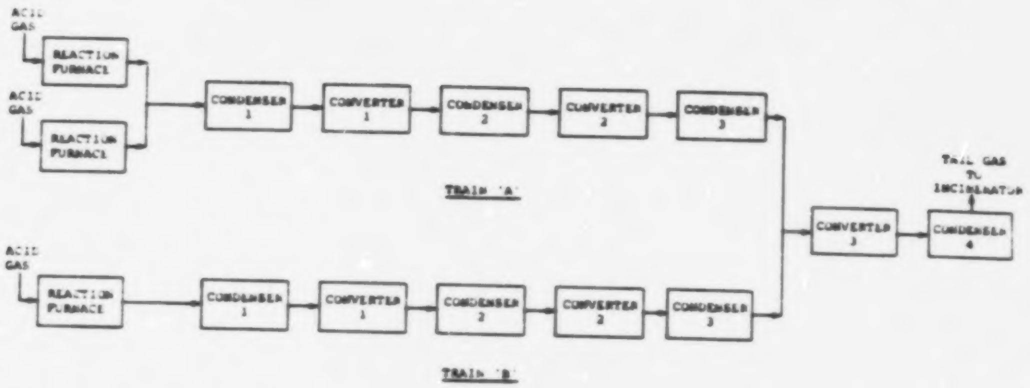
and that employed at the latter is the CBA. The schematics of the complete sulphur recovery plants at the above plants and at Taylor are shown in Figures 20(a)-(c).

CBA Process(43,52,62)

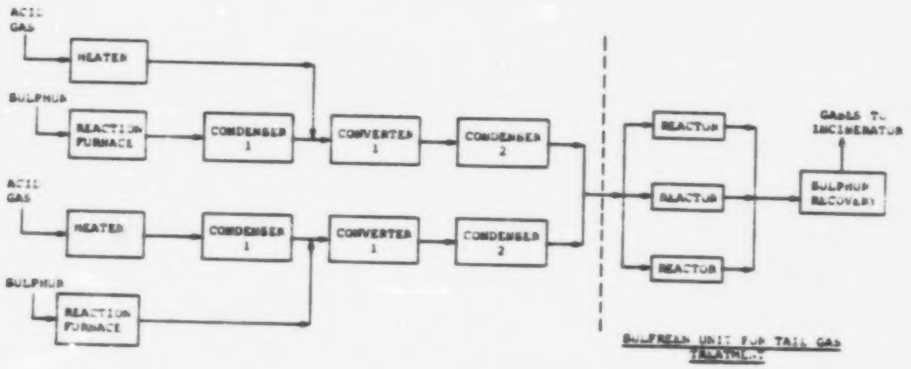
In this process developed by Amoco Production Co., shown schematically in Figure 21, the outlet gas from the final sulphur condenser is passed through a catalytic reactor maintained at about 126°-149°C, a temperature below the sulphur dew point. Such a low temperature favours the Claus Reaction (3) between H_2S and SO_2 to produce sulphur vapour which condenses on the catalyst bed. Such removal of sulphur vapour from the gaseous mixture shifts the reaction equilibrium towards higher conversion to sulphur. The outlet gas from the reactor is then incinerated before discharging to the atmosphere.

The catalyst used in the process is the same as conventional Claus catalyst. With the progress of the reaction, the catalyst activity gradually decreases by sulphur condensation. In practice, two catalytic reactors are used for continuous operation. While one of these is on adsorption cycle, the other reactor is being regenerated by a portion of the hot gas from the first reactor. As the sulphur is desorbed by the hot gas, the catalyst bed is reactivated and becomes ready for sulphur adsorption cycle. The hot gas from the regenerator is cooled to separate the sulphur as liquid in a condenser, and the gas is recycled to the first reactor.

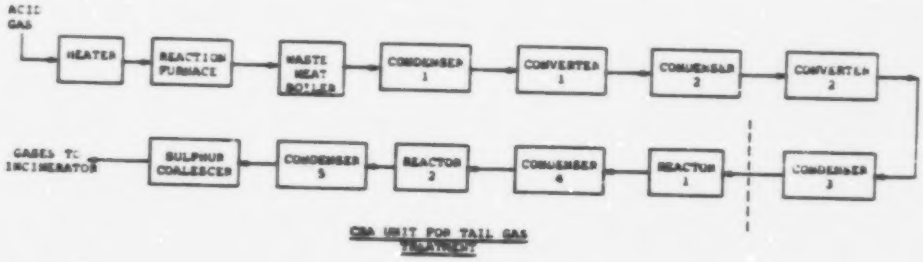
As the process is a continuation of the modified Claus system, it is suitable for retrofitting to an existing Claus plant. However, as the COS and CS_2 are not


ACID GAS
REACTION FURNACE
ACID GAS
REACTION FURNACE
CONDENSER 1
CONVERTER 1
CONDENSER 2
CONVERTER 2
CONDENSER 3
TRAIN 'A'
TAIL GAS TO INCINERATOR
CONVERTER 3
CONDENSER 4
ACID GAS
REACTION FURNACE
CONDENSER 1
CONVERTER 1
CONDENSER 2
CONVERTER 2
CONDENSER 3
TRAIN 'B'

(a): TAYLOR PLANT


ACID GAS
HEATER
SULPHUR
REACTION FURNACE
CONDENSER 1
CONVERTER 1
CONDENSER 2
ACID GAS
HEATER
CONDENSER 1
CONVERTER 1
CONDENSER 2
SULPHUR
REACTION FURNACE
REACTOR
REACTOR
REACTOR
GASES TO INCINERATOR
SULPHUR RECOVERY
SULFREEN UNIT FOR TAIL GAS TREATMENT

(b): FORT NELSON PLANT


ACID GAS
HEATER
REACTION FURNACE
WASTE HEAT BOILER
CONDENSER 1
CONVERTER 1
CONDENSER 2
CONVERTER 2
GASES TO INCINERATOR
SULPHUR COALESCER
CONDENSER 5
REACTOR 2
CONDENSER 4
REACTOR 1
CONDENSER 3
CBA UNIT FOR TAIL GAS TREATMENT

(c) PINE RIVER PLANT

FIGURE 20: SIMPLIFIED FLOW DIAGRAMS OF SULPHUR RECOVERY PLANTS IN BRITISH COLUMBIA

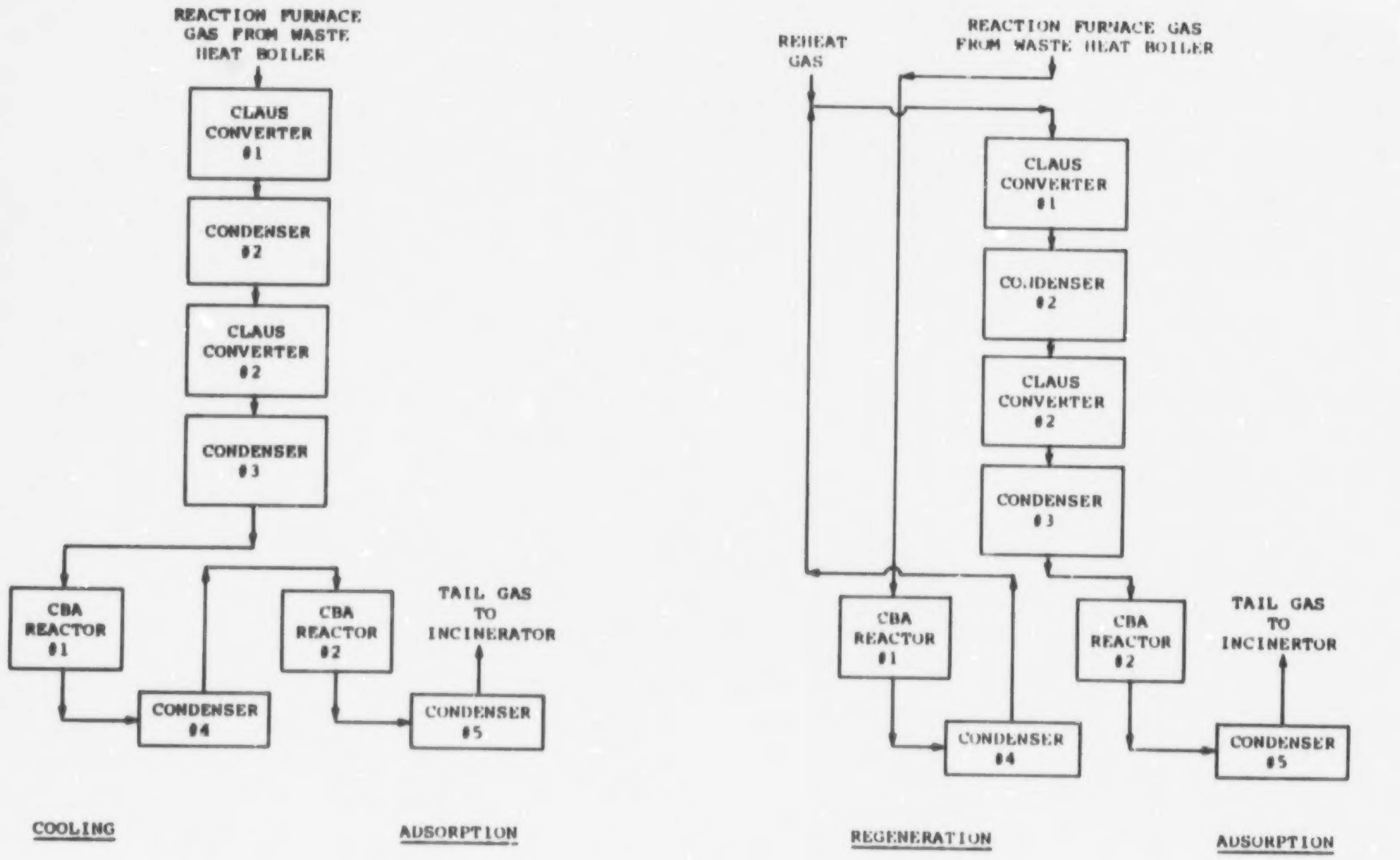

REACTION FURNACE GAS FROM WASTE HEAT BOILER
CLAUS CONVERTER #1
CONDENSER #2
CLAUS CONVERTER #2
CONDENSER #3
CBA REACTOR #1
CONDENSER #4
CBA REACTOR #2
TAIL GAS TO INCINERATOR
CONDENSER #5
COOLING
ADSORPTION
REHEAT GAS
REACTION FURNACE GAS FROM WASTE HEAT BOILER
CLAUS CONVERTER #1
CONDENSER #2
CLAUS CONVERTER #2
CONDENSER #3
CBA REACTOR #1
CONDENSER #4
CBA REACTOR #2
TAIL GAS TO INCINERTOR
CONDENSER #5
REGENERATION
ADSORPTION

FIGURE 21: SIMPLIFIED FLOW DIAGRAM OF AMOCO CBA PROCESS

reduced by the reaction, the main Claus plant operation should be controlled so that a minimum amount of these constituents are present in the tail gas. For maximum sulphur conversion, proper control of the stoichiometric $H_2S:SO_2$ ratio in the CBA reactor is essential. It has been reported that the overall sulphur recovery of 99.5% could be achieved with a two-stage CBA unit.

Sulfreen Process(44,46,52,62)

The process shown in Figure 22(a), was developed by Societe Nationale des Petroles d'Aquitaine (SNPA) of France and Lurgi Gessellschaft Gmbh of West Germany. It operates on the same principle as the CBA process and in the same temperature range but with an activated alumina catalyst. The major difference between the two processes lies in the catalyst regeneration method. In the Sulfreen process a closed-loop regeneration system by a heated inert gas is practised. The operating variables, sulphur recovery efficiency and control prerequisites are similar to that for the CBA process.

Simplified flow schematics of a few other commercially available tail gas treatment processes are shown in Figures 22(b) to (e).

6.1.3.3. Sulphur Recovery Efficiency and Monitoring

The recovery efficiency of a sulphur plant can be determined by the following methods:

1. $$\%R = \frac{S_p}{(S_p + S_e)} \times 100$$

2. $$\%R = \frac{(S_i - S_e)}{S_i} \times 100$$

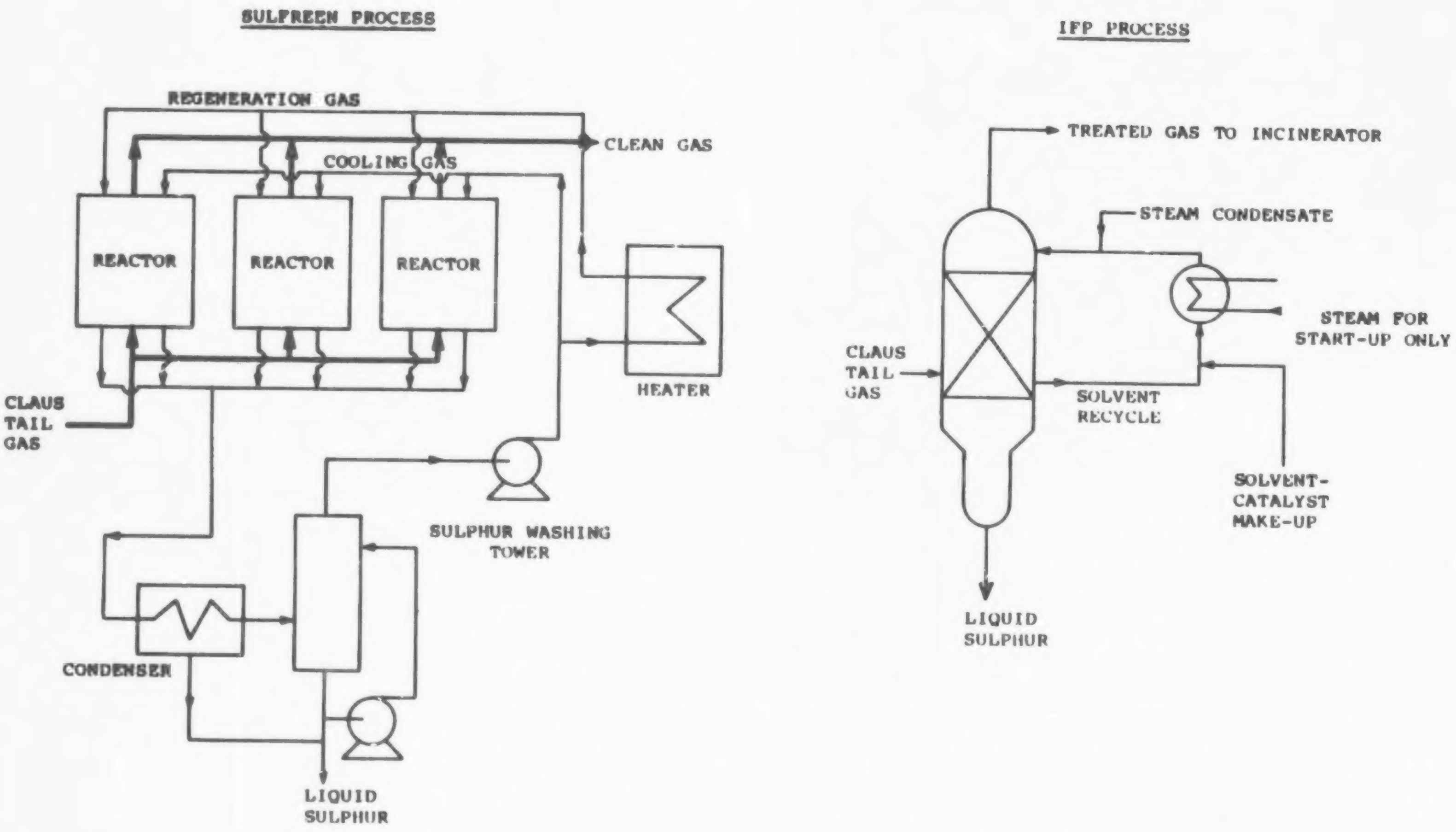

SULFREEN PROCESS
REGENERATION GAS
CLEAN GAS
COOLING GAS
REACTOR
REACTOR
REACTOR
HEATER
CLAUS TAIL GAS
SULPHUR WASHING TOWER
CONDENSER
LIQUID SULPHUR
IFP PROCESS
TREATED GAS TO INCINERATOR
STEAM CONDENSATE
STEAM FOR START-UP ONLY
CLAUS TAIL GAS
SOLVENT RECYCLE
SOLVENT-CATALYST MAKE-UP
LIQUID SULPHUR

FIGURE 22(a): FLOW SCHEMATICS OF TAIL GAS TREATMENT PROCESSES - SULFREEN AND IFP PROCESSES

SCOT PROCESS

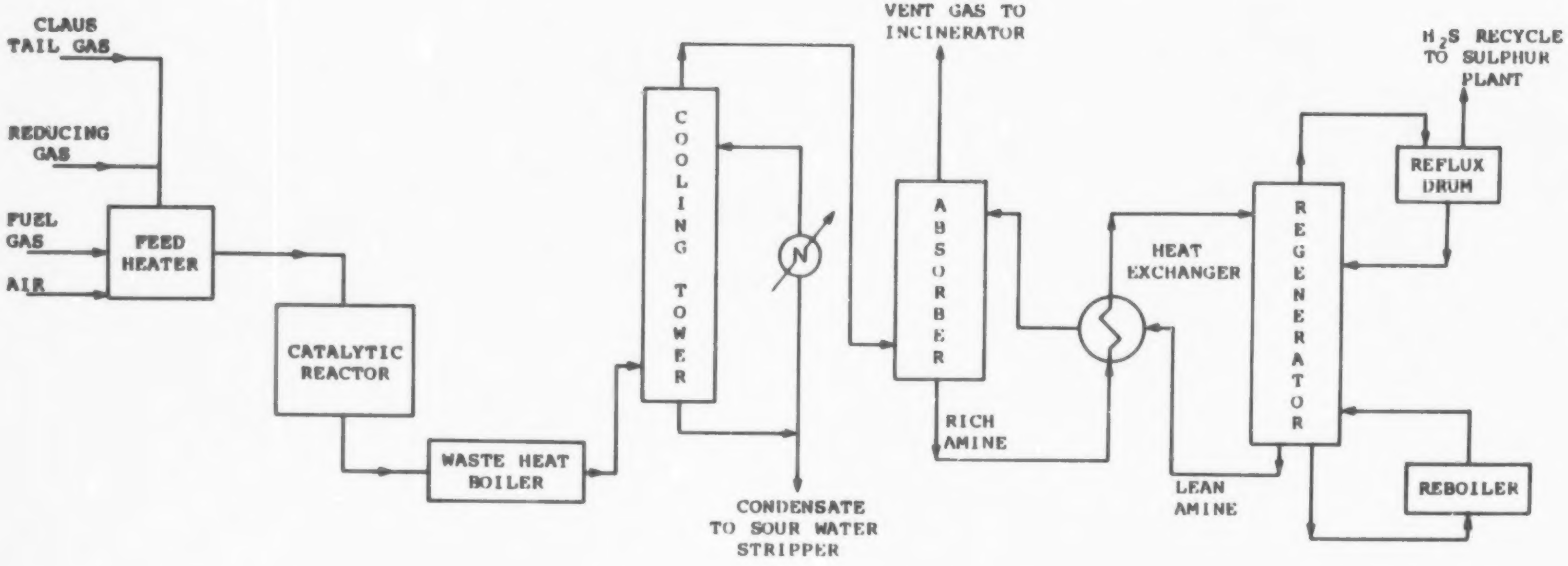

CLAUS TAIL GAS
REDUCING GAS
FUEL GAS
AIR
FEED HEATER
CATALYTIC REACTOR
WASTE HEAT BOILER
COOLING TOWER
CONDENSATE TO SOUR WATER STRIPPER
VENT GAS TO INCINERATOR
ABSORBER
RICH AMINE
HEAT EXCHANGER
LEAN AMINE
REGENERATOR
H_2S RECYCLE TO SULPHUR PLANT
REFLUX DRUM
REBOILER

FIGURE 22(b): FLOW SCHEMATICS OF TAIL GAS TREATMENT PROCESS - SCOT PROCESS

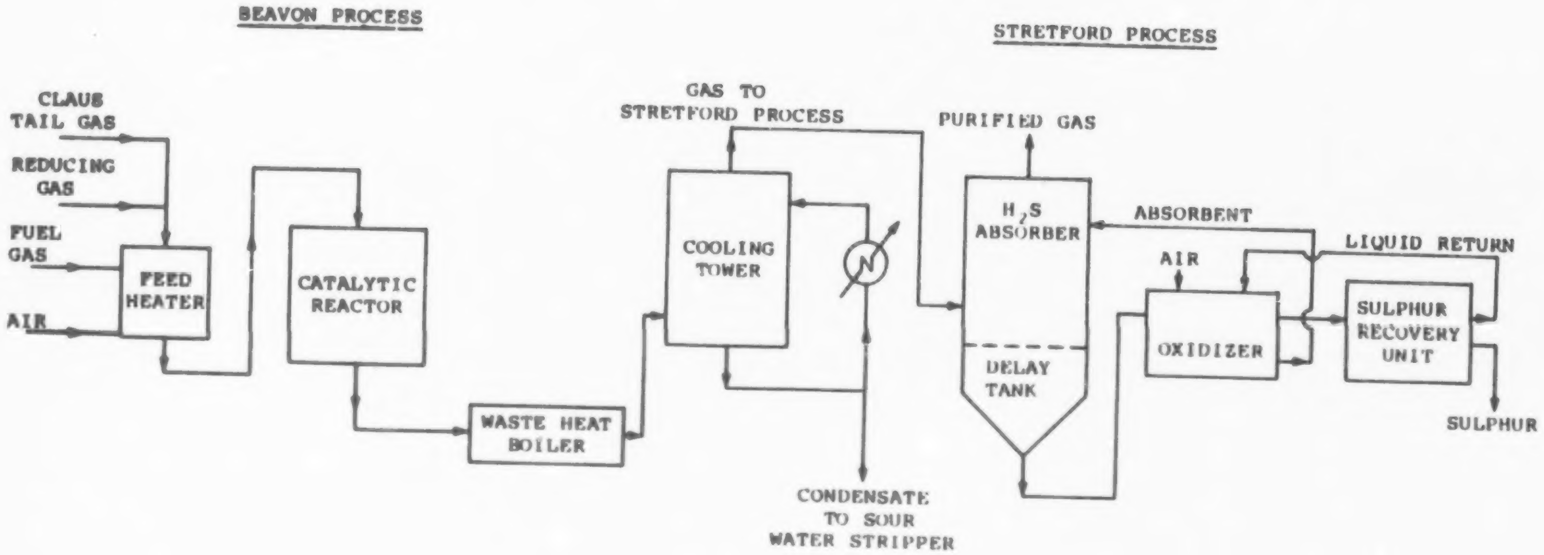

BEAVON PROCESS
STRETFORD PROCESS
CLAUS TAIL GAS
REDUCING GAS
FUEL GAS
AIR
FEED HEATER
CATALYTIC REACTOR
WASTE HEAT BOILER
GAS TO STRETFORD PROCESS
COOLING TOWER
CONDENSATE TO SOUR WATER STRIPPER
PURIFIED GAS
H_2S ABSORBER
DELAY TANK
ABSORBENT
AIR
OXIDIZER
LIQUID RETURN
SULPHUR RECOVERY UNIT
SULPHUR

FIGURE 22 (c): FLOW SCHEMATICS OF TAIL GAS TREATMENT PROCESS - BEAVON-STRETFORD PROCESS

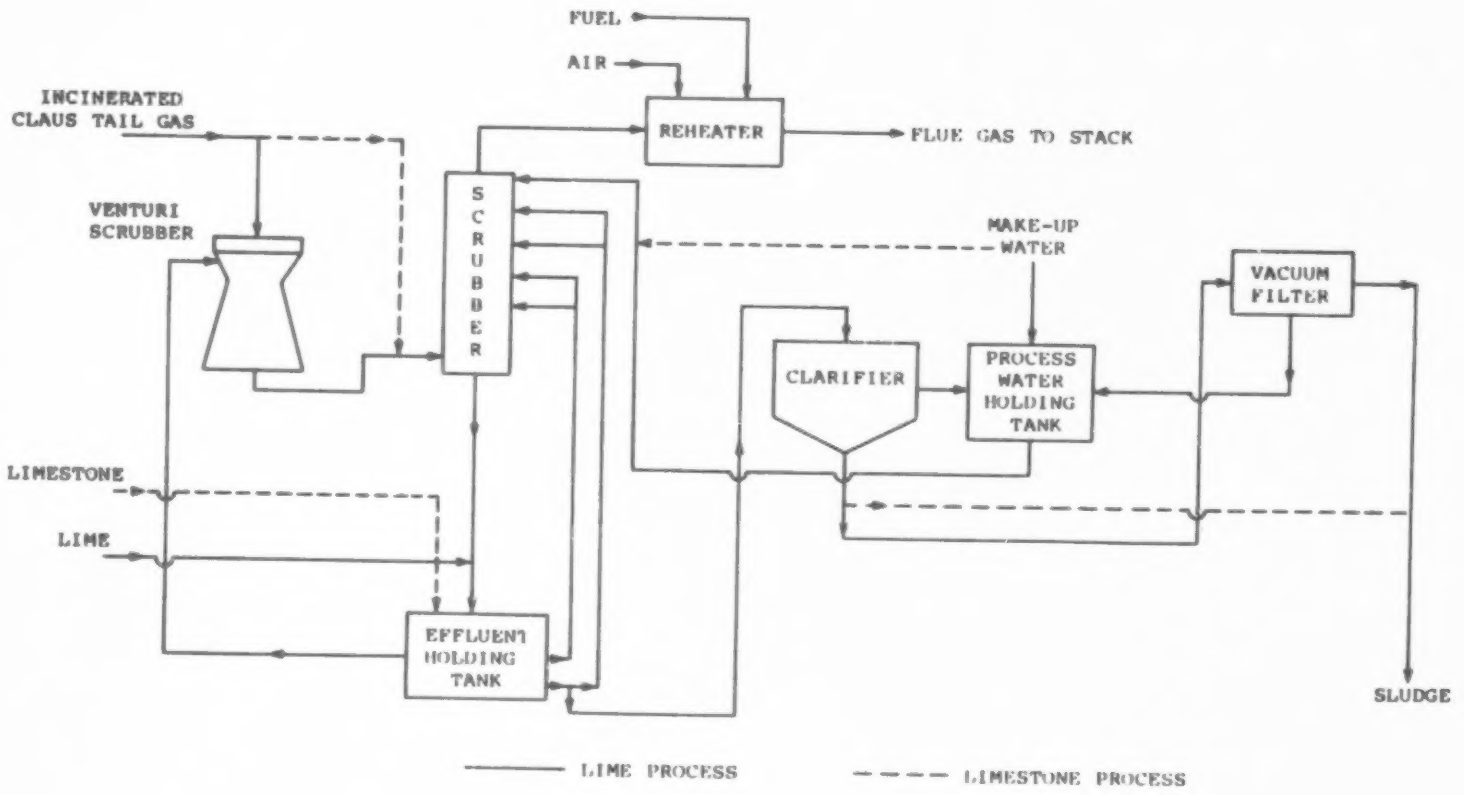

FUEL
AIR
INCINERATED CLAUS TAIL GAS
REHEATER
FLUE GAS TO STACK
VENTURI SCRUBBER
SCRUBBER
MAKE-UP WATER
VACUUM FILTER
CLARIFIER
PROCESS WATER HOLDING TANK
LIMESTONE
LIME
EFFLUENT HOLDING TANK
SLUDGE
LIME PROCESS
LIMESTONE PROCESS

FIGURE 22(d): FLOW SCHEMATICS OF TAIL GAS TREATMENT PROCESS - LIME/LIMESTONE PROCESS

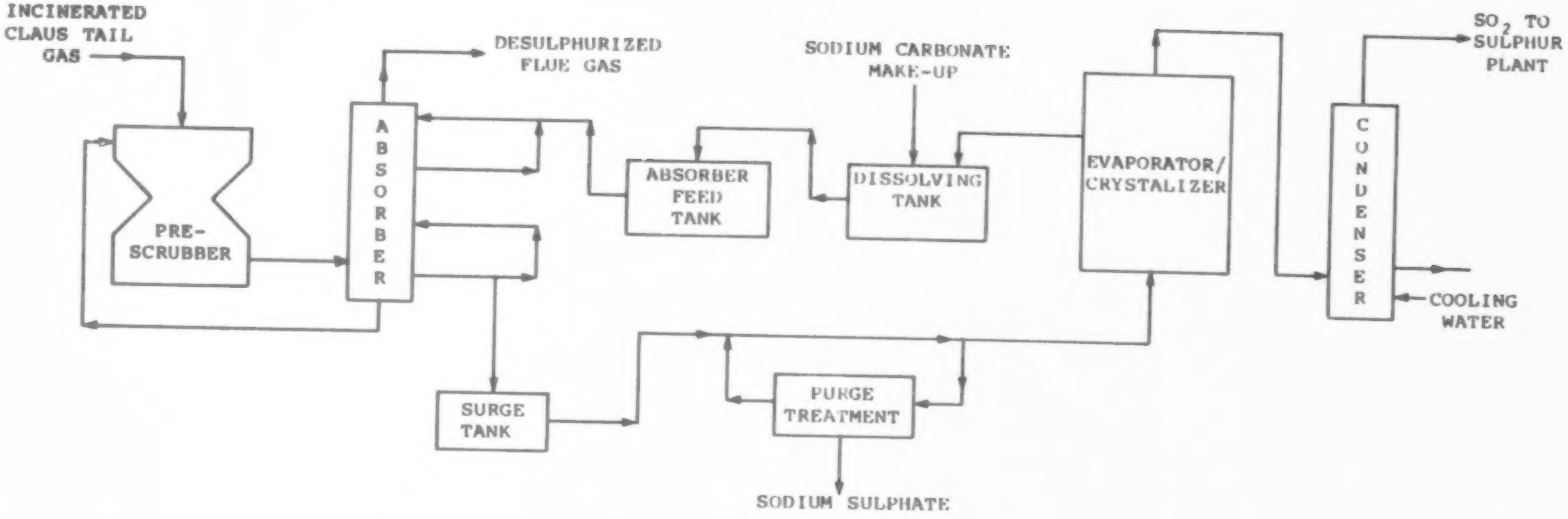

INCINERATED CLAUS TAIL GAS
PRE-SCRUBBER
ABSORBER
DESULPHURIZED FLUE GAS
ABSORBER FEED TANK
SODIUM CARBONATE MAKE-UP
DISSOLVING TANK
EVAPORATOR/ CRYSTALIZER
CONDENSER
SO_2 TO SULPHUR PLANT
COOLING WATER
SURGE TANK
PURGE TREATMENT
SODIUM SULPHATE

FIGURE 22 (e): FLOW SCHEMATICS OF TAIL GAS TREATMENT PROCESS - WELLMAN-LORD PROCESS

3. $$\%R = \frac{S_p}{S_i} \times 100$$

where R = sulphur recovery
S_p = sulphur as product
S_e = sulphur emitted through the sulphur plant incinerator stack
S_i = sulphur in acid gases to the sulphur plant

However, to determine an overall recovery efficiency as well as to control total emission of sulphurous gases from the gas processing plant, the following method seems to be more appropriate:

4. $$\%R = \frac{S_p}{S_{In}} \times 100$$

where S_{In} represents the total sulphur entering the plant with the raw gas.

The determination of raw inlet-gas sulphur content in a continuous manner is often difficult, because of the variations in both the flow rates of the gas and its sulphur content. The technique for gas flow rate measurement and the analytical method for sulphur content must be reliable and accurate. The wet chemical Tutweiler method for sulphur analysis is potentially accurate to ±1%, and has been in use for a long time. However, it can be used as a spot check only. The gas chromatographic method is capable of analysing all sulphur species separately and is very accurate. As the gas samples, especially with low H_2S concentrations, deteriorate with time, the analysis should be done immediately(34).

The sulphur production rate is measured by a gauge in the pit for holding liquid sulphur. When the sulphur volume is measured, appropriate temperature correction should be applied. Usually with proper care the sulphur production could be made quite accurately.

In practice, at the large plants the Claus plant tail gas is monitored continuously for H_2S/SO_2 ratio to optimize air supply to the reaction furnace. With the introduction of modern analytical tools, such as gas chromatograph, ultra-violet and infra-red monitors, and computers, optimization of air supply based on continuous H_2S and SO_2 analysis has become a part of the Claus process control system. It is also a regular practice to monitor the tail gas incinerator stack continuously for the volume, temperature and SO_2 content of the flue gas. Some monitoring systems also provide integrated mass emission rate of sulphur on a continuous basis, so that the sulphur recovery efficiency could be determined as necessary. At some plants the tail gas is analysed for H_2S, COS, CS_2 and SO_2 prior to its incineration(22). However, such a system fails to account for any possible loss of sulphur as entrained vapour.

In addition to the above monitoring of sulphur compounds, the regulatory agencies often require complete sulphur balance at large plants to ascertain compliance with the total sulphur emission rates. To carry out such a material balance it is necessary to determine the following:

1. Sulphur in the plant inlet gas
2. Sulphur in the fuel gas used
3. Sulphur in the condensates
4. Sulphur in the flared gas
5. Sulphur produced
6. Sulphur emission through the incinerator stack.

If there is no unaccounted for loss of sulphur, the item 1 should equal to the sum of the items 2 to 6 inclusive. In most cases, the sulphur in the fuel gas and condensates would be negligible. However, any information on these items would be useful in determining the overall sulphur recovery efficiency at a plant if S_{In} in method 4 is substituted by the following:

$$5. \quad \%R = \frac{Sp}{S_p + S_f + S_c + S_{fl} + S_e} \times 100$$

where S_f = sulphur in the fuel gas
S_c = sulphur in the condensate
S_{fl} = sulphur in the flare

The inclusion of S_f, S_c and S_{fl} in the above method of determination of sulphur recovery efficiency provides the plant operators with an opportunity to rectify any problems with the respective processing units and to control avoidable sulphur emissions from these sources. To control unnecessary flaring of raw or acid gases upstream of the sulphur plant, an allowance in terms of percentage of sulphur recovery could be stipulated by the regulatory authorities to cope with start-ups, shut-downs or abnormal operating conditions.

It should be noted that with the currently available analytical instruments and computerized process control systems, a sulphur recovery plant of the Claus type could be operated efficiently without any excessive emission of sulphur compounds. The sulphur recovery efficiency of such a plant could be maintained within 0.1% of the designed figure in a plant processing an inlet gas with more than 10 t/d of sulphur and an acid gas containing more than 40% H_2S(64).

6.1.3.4. Proposed Guidelines For Sulphur Recovery Plant Emission

It is proposed that any revision to the emission objectives for sulphur recovery plants should be based on the information provided above. The following relevant aspects should be considered:

1. The present requirement of sulphur recovery for plants processing inlet gases containing more than 10 t/d of sulphur should be retained. However, if environmental considerations dictate, such a requirement could also be applied to plants smaller than the above size.

2. The level of sulphur recovery efficiency should be stipulated on the basis of H_2S content of the acid gas which appears to be the major governing factor. Any excessive CO_2 in the raw gas could be removed by selective absorption at the desulphurization stage.

3. The minimum sulphur recovery level could be stipulated to the nearest 0.1%, as the presently available monitoring and control systems are able to maintain such a level under normal operating conditions.

4. An allowance for the recovery level on a monthly or quarterly average basis could be stipulated to accommodate flaring of raw or acid gas during start-up, shut-down and emergency conditions. The province of Alberta allows a quarterly average recovery of 0.3% less than the stipulated recovery level for such exigencies(64).

5. In the procedure for determination of sulphur recovery efficiency the quantity of sulphur present in the condensates produced should also be included.

6.2 Liquid Effluents

The potential sources of aqueous wastes in a typical gas processing plant are shown schematically in Figure 23. Major characteristics of these effluents and the applicable control technologies are summerized in Table 8. It should be noted that if any cooling of the plant water is practised in a tower, another waste stream of cooling tower blowdown would require handling. However, in British Columbia none of the processing plants has any cooling tower, and this effluent will not be discussed.

The actual number of effluent streams and their characteristics would depend upon the composition of the raw inlet gas, the processing configuration, and the operating efficiency of the processing units. If the raw gas processed is relatively sweet and no sulphur is recovered, both the quantity and quality of effluents generated are expected to be of less concern and may need simpler treatment prior to disposal. Typical plant effluents can be classified into four following groups:

1. Sanitary effluent
2. Storm water and plant area run-off
3. Water treatment and boiler blowdown
4. Process wastewater

At well-head and smaller installations the sanitary effluent is conventionally treated and any process wastewater is disposed of by either deep-well injection or incineration in the flare system. There are

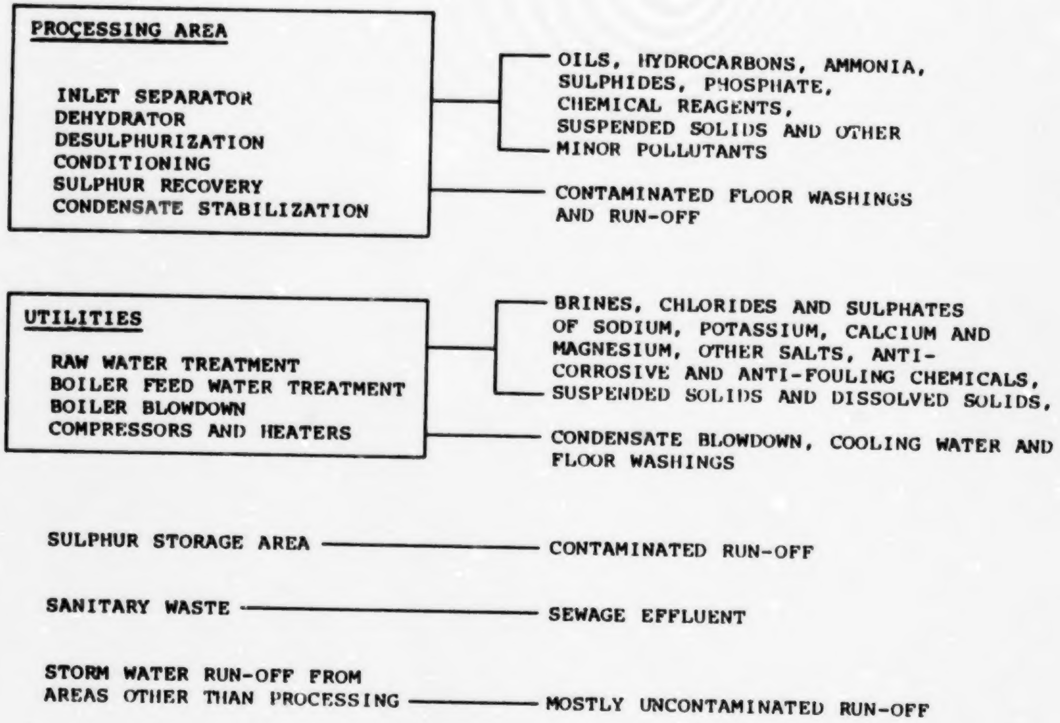

PROCESSING AREA
INLET SEPARATOR
DEHYDRATOR
DESULPHURIZATION
CONDITIONING
SULPHUR RECOVERY
CONDENSATE STABILIZATION
OILS, HYDROCARBONS, AMMONIA, SULPHIDES, PHOSPHATE, CHEMICAL REAGENTS, SUSPENDED SOLIDS AND OTHER MINOR POLLUTANTS
CONTAMINATED FLOOR WASHINGS AND RUN-OFF
UTILITIES
RAW WATER TREATMENT
BOILER FEED WATER TREATMENT
BOILER BLOWDOWN
COMPRESSORS AND HEATERS
BRINES, CHLORIDES AND SULPHATES OF SODIUM, POTASSIUM, CALCIUM AND MAGNESIUM, OTHER SALTS, ANTI-CORROSIVE AND ANTI-FOULING CHEMICALS, SUSPENDED SOLIDS AND DISSOLVED SOLIDS,
CONDENSATE BLOWDOWN, COOLING WATER AND FLOOR WASHINGS
SULPHUR STORAGE AREA
CONTAMINATED RUN-OFF
SANITARY WASTE
SEWAGE EFFLUENT
STORM WATER RUN-OFF FROM AREAS OTHER THAN PROCESSING
MOSTLY UNCONTAMINATED RUN-OFF

FIGURE 23: POTENTIAL SOURCES OF EFFLUENT GENERATION FROM A NATURAL GAS PROCESSING FACILITY

effluent treatment facilities at the major plants in British Columbia. While the plants at Fort Nelson and Pine River have individual systems, the effluents from the Taylor plant are sent to the adjacent oil refinery for combined treatment with other wastewaters.

6.2.1 Sanitary Effluent

Sanitary wastewater from various areas of the plant is usually collected in local sumps and then combined for treatment. This effluent can be treated by conventional primary and secondary treatment methods. Tertiary treatment and/or disinfection of the final effluent may be required sometimes if the treated effluent is discharged into any waterbody or on land with high watertable.

Depending on their compatibility, the sanitary effluent can be treated by conventional methods with other raw or partially treated wastewater. For example, at Fort Nelson plant the processing area effluents, except those from water treatment and boiler blowdown, are treated jointly with the sanitary wastewater. At the Pine River plant, however, only the floor washings from the utility and shop buildings are treated with the sanitary effluent. The available methods for treatment of the former type mixed effluent are discussed in a later section.

At the Pine River plant the sewage effluent from local sumps is sent to the treatment unit by grinder type pumps. The floor washings mentioned above are first passed through an oil-water separator to reduce the oil and grease content, and then to the treatment unit which is a packaged extended aeration unit. The total volume of the influent to the treatment unit is averaged at 5 m^3/d with a maximum oil and grease content of 10 mg/L. The treated effluent is discharged to evaporation ponds where a long residence time tends to improve the effluent quality(65). It appears that the subject treatment methods are adequate and produce a reasonable quality effluent in the ponds. Typical analysis of the pond water is shown in Table 20.

6.2.2. Storm Water and Plant Area Run-Off

The surface run-off from all processing and by-product storage areas may become contaminated and require treatment prior to its discharge. The storm water from the peripheral land should be diverted away from the above areas and allowed to follow the natural drainage pattern of the land.

The contaminated surface runoff water may contain various amounts of treating chemicals, oily substances and acidic leachates from sulphur storage areas. Good housekeeping practice is essential in minimizing contamination by chemicals and oils. However, the drainage water from sulphur storage areas is often characterized by low pH and high metal content. Several possible means of sulphuric acid production, by both chemical and microbial reactions, from elemental sulphur and other sulphur compounds present in it have been suggested in the literature(67-70). The acidity of the drainage water from sulphur storage area would vary from site to site, but in some cases pH values of 1 and 2 have been reported(71). Such acidic water dissolves metals from soil, especially aluminum and iron, and transforms the soil unfit for vegetation. If this type of drainage water is discharged into any surface water, the dissolved metals may affect the aquatic life adversely. In British Columbia, Petrosul International Ltd. operates

TABLE 20

TYPICAL ANALYSIS OF EVAPORATION POND WATER AT PINE RIVER PLANT

(Sewage and Floor Washings Effluents)

Constituent	Concentration
Total Suspended Solids	40
Total Dissolved Solids	340
B.O.D.$_5$	74
Alkalinity (as $CaCO_3$)	212
Chemical Oxygen Demand	360
Calcium	42
Magnesium	8.4
Sodium	54
Boron	0.16
Chloride	22
Sulphate	95
Phosphate Total (as P)	3.68
Nitrite (as N)	0.003
Nitrate (as N)	1.13
Ammonia (as N)	0.07
Total Kjeldahl Nitrogen	16.0
Oil and Grease	1
Faecal Coliform	Less than 4*

concentrations are in mg/L unless otherwise stated

* Per 100 mL

Source: (66)

sulphur pelletizating plants at the sites of sulphur storage, and treats the area run-off effluent for final pH between 6.5 and 8.5.

The general practice in the industry is to collect this water and neutralize to a safer pH and metal contents before discharging to land or any water body. However, some possible measures(71) to prevent acid generation in the storage area can be undertaken as follows:

- a high ground with low water table and impervious soil should be chosen for sulphur storage;
- the natural drainage should be prevented from entering the area;
- the base of the storage area must be firm and designed to drain away any water;
- if possible, minimize storage quantity and provide a roof or waterproof cover over the stored material.

6.2.3 Water Treatment and Boiler Blowdown

A number of chemicals are used to treat raw and boiler feed water as well as for internal cleaning of boiler to control softness, solids content, pH, corrosion and scale formation in the boiler tubes. Some of the commonly used chemicals for these purposes are listed in Table 21. The characteristics of the effluents would depend on the types of chemicals used to treat the raw intake water.

The effluent from raw water treatment system consists of the clarifier filter backwash, blowdowns from softeners and the wash water from demineralizers. The filter backwash contains mostly soil-derived solids and traces of coagulants. As lime and/or soda ash are used as softeners, these chemicals are normally present in the

TABLE 21

CHEMICAL ADDITIVES COMMONLY USED IN BOILER WATER TREATMENT AND BOILER CLEANING

Control Purpose	Chemical Additive
Softening and Demineralization	Lime (quicklime, hydrated lime), Salt, Soda ash, Ion exchange material
Clarification	Alum, Sodium aluminate, Ferric chloride, Ferrous sulphate and Coagulation Aids (bentonite, kaolin, activated alumina, carbon, polyelectrolytes)
Scale and Solids Deposition	Alginates, Di-and tri-sodium phosphates, Ethylene diaminetetra-acetic acid (EDTA), Nitrilotriacetic acid (NTA), Lignins, Tannins, Starch, Polyacrylamides, Polyacrylates, Polymethacrylates, Phosphates
Corrosion	Sodium sulphite, hydrazine, Chromates, Polyphosphates, Volatile or filming Amines, Morpholine
pH	Sodium hydroxide, Sodium carbonate, Ammonia, Lime, Hydrochloric acid, Sulphuric acid

Source: (72)

blowdown from these units. High concentrations of primary chlorides and sulphates of sodium, potassium, calcium and magnesium, and other salts are expected in the wastewater from ion-exchange demineralizers.

To maintain the desired solids concentration in the boiler steam drum, typically around 2 000 mg/L, boiler blowdown is practised either intermittently or continuously. This blowdown volume can be significant, and contains traces of chemicals used in feed water treatment and for internal cleaning of the boiler tubes. It may also contain some metals, such as iron, copper and zinc, as the products of boiler corrosion. Some proprietary chemicals used may contribute copper, chromium and phenols to the boiler blowdown, and their use should be avoided. Certain cleaning chemicals present in the blowdown may contribute towards high dissolved solids and significant oxygen demand. Usually, however, the boiler blowdown is of high quality and a major portion of it can be recovered for reuse at the plant. In British Columbia the plant at Pine River utilizes the waste heat to recover about 90% of the boiler blowdown for use in the steam system.

As far as the disposal of these two effluents is concerned, the general practice in the industry is to send both to a holding lagoon or evaporation pond. Only treatment which may be required is pH adjustment, as the pH would depend on whether acidic or alkaline chemicals are used. Typical analysis of this combined effluent in the pond from the Pine River plant is shown in Table 22. The presence of the trace amount of oil is probably from the leaks in the seals and pumps carried into the pond by floor washings. Final disposal of the pond effluent would depend on the location. At the Fort Nelson plant this effluent is discharged into a river, but at the Pine River an evaporation pond is used. After a pilot test

TABLE 22

TYPICAL ANALYSIS OF EVAPORATION POND WATER AT PINE RIVER PLANT

(Boiler Blowdown, Filter Backwash and Boiler Cleaning Effluents)

Constituent	Concentration
Total Suspended Solids	<1
Total Dissolved Solids	350
B.O.D.$_5$	<10
Alkalinity (as $CaCO_3$)	24
Chemical Oxygen Demand	43
Calcium	56
Magnesium	11
Sodium	30
Boron	0.01
Chloride	30
Sulphate	160
Phosphate Total (as P)	0.03
Nitrite (as N)	0.002
Nitrate (as N)	0.13
Ammonia (as N)	<0.01
Total Kjeldahl Nitrogen	0.43
Oil and Grease	2

concentrations are in mg/L

Source: (66)

program at the latter plant a portion of this effluent is sprayed on land during summer months$^{(66)}$.

6.2.4 Process Wastewater

The process wastewater from a typical gas processing plant originates from the following sources:

1. Sour water from the inlet gas scrubber, regenerator reflux accumulators and other vessels (mostly sulphides and condensates).

2. Acid gas removal (mostly suspended solids, treating chemicals and hydrocarbon oils).

3. Gas dehydration (residual chemicals and condensates).

4. Floor washings from various processing areas (hydrocarbon oils and treating chemicals).

There are several alternatives available for treatment and disposal of the above effluents, and the choice would depend on the particular circumstances and the regulations on discharge effluent quality. Broadly, the major alternatives are as follows:

1. Deep well injection of the untreated or partially treated effluent.

2. Segregation of wastewaters, recovery of chemicals and incineration of the residues.

3. Treatment of individual effluents or of combined effluent with other wastewaters.

Although the disposal in deep wells has been practised for some time, recently this method is not encouraged by the regulatory agencies because of potential environmental risks. This aspect is discussed later.

With proper planning and use of closed drains and local sumps, individual wastewaters can be handled adequately to prevent any positive discharge. At the Pine River plant such a system, shown schematically in Figure 24, is used without any apparent problems. The effluents from the inlet separators and fuel gas scrubbers are sent to a low pressure flare knock-out drum. The volatile substances are driven off from the effluent by heating and sent to the flare, while the liquid portion is pumped to a holding tank. Any liquid collected in the high pressure flare knock-out drum is also sent to the above holding tank after separation of the volatiles by heating.

The washings from the processing areas and equipment which may contain any sulfinol solution is pumped to the same holding tank. The combined effluent in this tank contains various proportions of glycol, methanol, diesel oil, sulfinol and water. An auxiliary burner in the tail gas incinerator is used to burn this effluent, which is injected intermittently at a maximum rate of 0.00378 m^3/min.

As the inlet gas to the plant is dry and the flare knock-out drums trap small amounts of liquids during normal operation, the primary source of wastewater for incineration is the processing areas. The total volume of the effluent is such that only occasional incineration is required. To cope with any upset conditions any overflow of liquid from the holding tank is directed to an

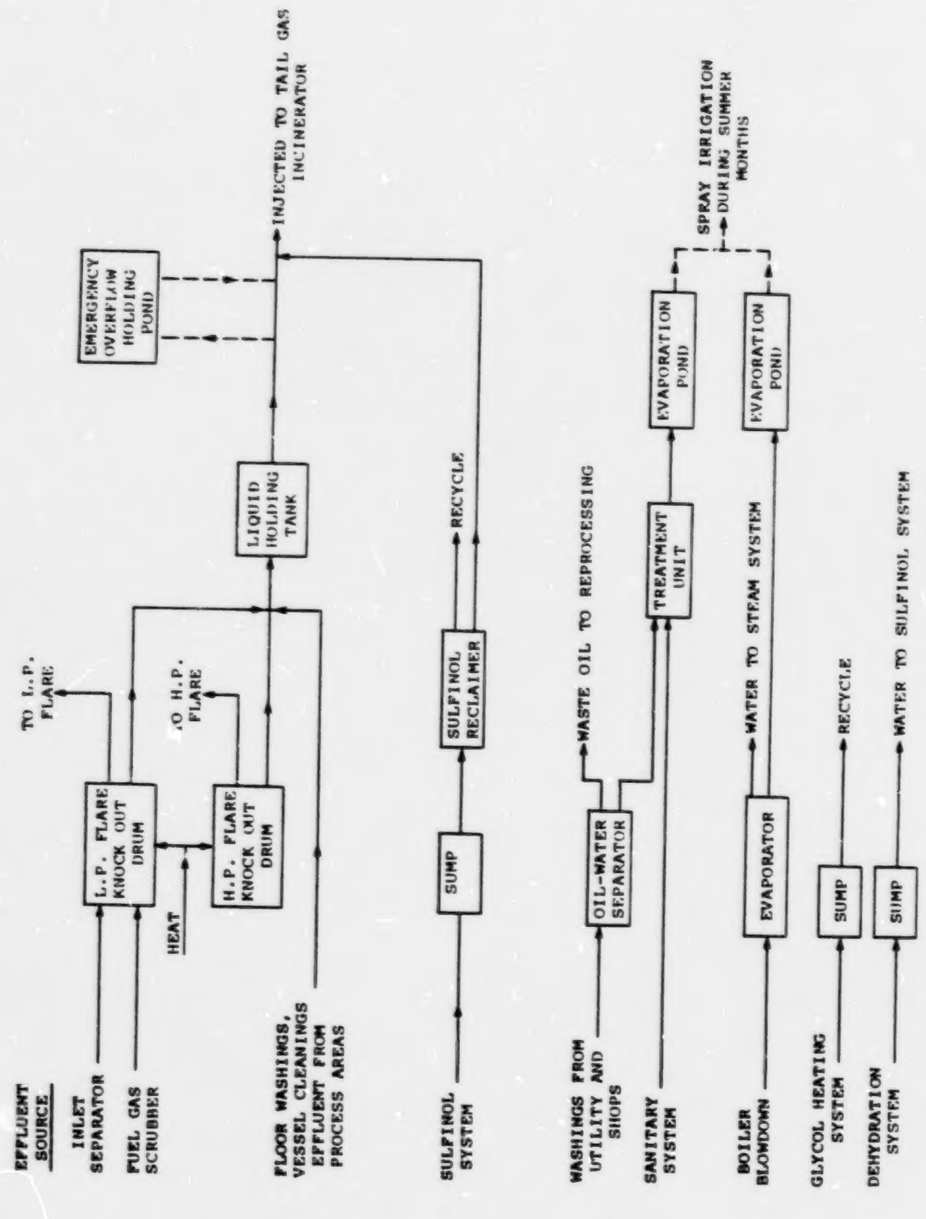

EFFLUENT SOURCE
INLET SEPARATOR
FUEL GAS SCRUBBER
L.P. FLARE KNOCK OUT DRUM
HEAT
H.P. FLARE KNOCK OUT DRUM
TO L.P. FLARE
TO H.P. FLARE
FLOOR WASHINGS, VESSEL CLEANINGS EFFLUENT FROM PROCESS AREAS
LIQUID HOLDING TANK
EMERGENCY OVERFLOW HOLDING POND
INJECTED TO TAIL GAS INCINERATOR
SULFINOL SYSTEM
SUMP
SULFINOL RECLAIMER
RECYCLE
WASHINGS FROM UTILITY AND SHOPS
OIL-WATER SEPARATOR
WASTE OIL TO REPROCESSING
SANITARY SYSTEM
TREATMENT UNIT
EVAPORATION POND
EVAPORATION POND
SPRAY IRRIGATION DURING SUMMER MONTHS
BOILER BLOWDOWN
EVAPORATOR
WATER TO STEAM SYSTEM
GLYCOL HEATING SYSTEM
SUMP
RECYCLE
DEHYDRATION SYSTEM
SUMP
WATER TO SULFINOL SYSTEM

FIGURE 24: FLOW SCHEMATICS OF EFFLUENT TREATMENT SYSTEMS AT PINE RIVER PLANT

emergency holding pond where it is held until the effluent is returned to the tank for incineration. This emergency pond is clay-lined to minimize any seepage of the effluent into the ground.

Closed drains and sumps are used to collect the liquids in the major chemical handling areas, namely the sulfinol and glycol systems. The chemicals collected are mostly recycled to the respective system. The water from the dehydrators is reused in the sulfinol system. After its use a portion of the sulfinol solution is generally degraded which requires separation. A reclaimer is used to separate the degradation product, typical composition of which is given in Table 23. This degradation product is disposed of by burning in the tail gas incinerator through an auxiliary burner at a maximum rate of 30 kg/h.

The third alternative mentioned above is generally practised where the wastewater volumes are large and the final effluent is discharged into a surface waterbody. There are several treatment methods available for the process wastewater which are discussed in the following section.

6.2.4.1 Process Wastewater Treatment

The effluents from the processing areas can be treated individually or jointly with the sanitary wastewater by employing a combination of chemical, physical and biological treatment methods listed in Table 24.

The major criteria for selection of any method would be to remove the organic chemicals from the effluent. The most common method is one of the biological oxidation processes, such as extended aeration pond and

TABLE 23

DEGRADATION PRODUCT OF SULFINOL SOLUTION

Simplified Chemical Formula	Percent (weight/weight)
$C_7 H_{13} NO_3$	60
$C_7 H_{13} NSO_2$	15
$C_7 H_{13} NS_2O$	15
$C_6 H_{15} NO_2$	5
$C_4 H_8 SO_2$	5

Source: (65)

TABLE 24

TREATMENT PROCESSES APPLICABLE TO EFFLUENTS FROM NATURAL GAS PROCESSING PLANTS

Process	Purpose
Storage and equalization pond	Minimize shock waste load and provide system stability
Screening	Removal of floatable solids and grit
Oil separator	Oil and grease removal
Stripping	Remove ammonia and hydrogen sulphide
Neutralization	Attain the desired pH of the effluent
Chemical coagulation and sedimentation	Remove suspended solids
Biological treatment	Remove biodegradable organics
Filtration	Remove additional suspended solids
Carbon adsorption	Remove additional dissolved organics, colour and toxic substances
Ion-exchange	Remove dissolved inorganic solids

activated sludge. However, to prepare the wastewater for biological treatment it is necessary to reduce its H_2S, oily substances and suspended solids contents to suitable levels.

Hydrogen sulphide containing wastewater streams, the principal one of which is the inlet separator sour water, are usually passed through a stripper to reduce the H_2S level to approximately 10 mg/L. There are several designs of sour water stripper using trayed column and a stripping medium of steam, natural gas or flue gas. A typical stripper, shown in Figure 25, is operated at a slightly reduced pressure and in the temperature range of 104° - 110°C with trays numbering from six to twelve. The efficiency of this type of stripper normally ranges from 96 to 99.5%, and the H_2S content of the stripped effluent could be reduced to 10 mg/L or less when the stripping is carried out at a low pH(73-76). The stripped H_2S is routed through the amine unit for eventual sulphur recovery, or can be flared if no sulphur is recovered.

The effluents containing oily substances are passed through an API oil-water separator where free oil, suspended solids and water are separated by gravity action. Free oil which rises to the surface of the water is skimmed off and some of the suspended materials sink to the bottom. The effluent water from the separator would still contain some oil and suspended solids, as the efficiency of an API separator varies from 60 to 99% depending on the particular design (76).

An API separator does not remove emulsified oil. If the oil content of the separator effluent is high enough to interfere with the downstream biological treatment, additional reduction of oil would be necessary. Any excessive amount of oil in the wastewater may affect

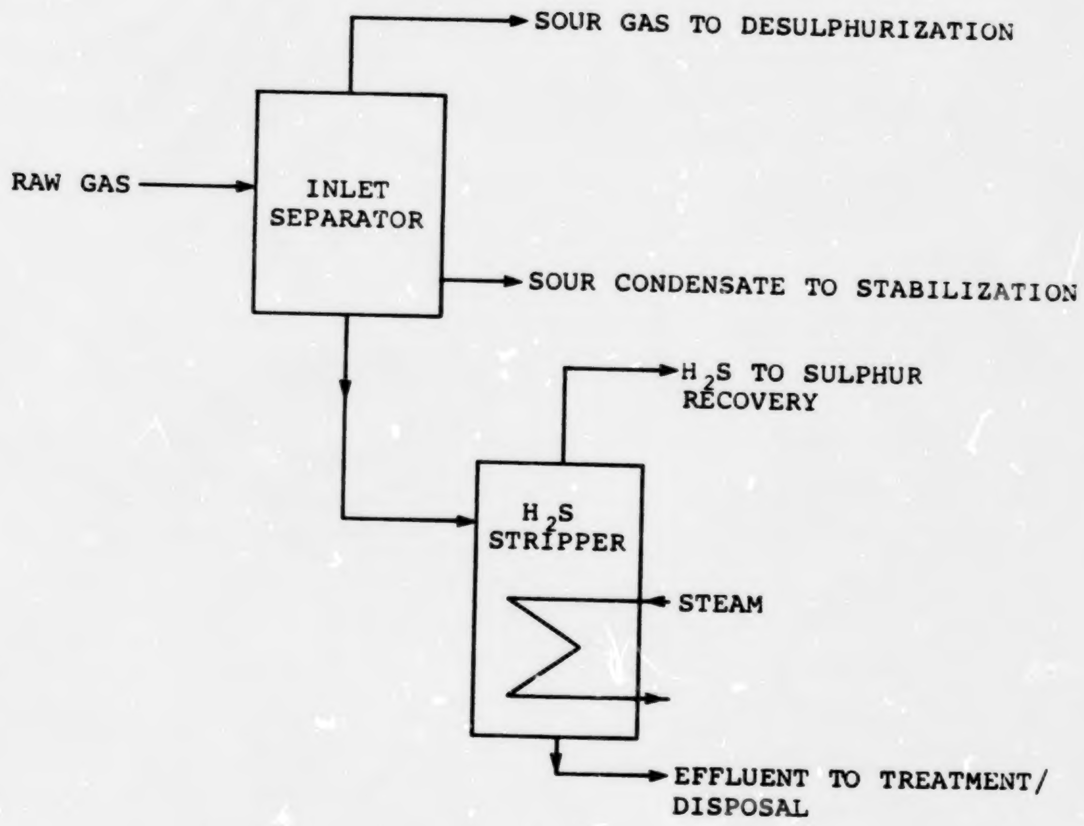

SOUR GAS TO DESULPHURIZATION
RAW GAS
INLET SEPARATOR
SOUR CONDENSATE TO STABILIZATION
H_2S TO SULPHUR RECOVERY
H_2S STRIPPER
STEAM
EFFLUENT TO TREATMENT/ DISPOSAL

FIGURE 25: TYPICAL FLOW DIAGRAM OF A SOUR WATER STRIPPER

the oxygen transfer rate during aeration. It may also increase the suspended solids content in the treated effluent by being adsorbed on the solid particles and thus lowering the specific gravity of the latter(76).

Additional oil is removed from the gravity separator effluent by dissolved air flotation, with or without any flocculant, in a number of oil refineries(75). However, to achieve effective separation of emulsified oil, a flocculating agent, such as alum or lime, is necessary. At the Waterton gas processing plant in Alberta, the emulsified oil is removed in a rapid-mix tank by flocculation with lime. A clarifier is used to separate the lime floc and adsorbed oil from the water(77).

The effluent after oil removal contains mostly organic chemicals used in the gas treatment processes and organic matter from the sewage system if the latter is combined upstream. Biological treatment methods generally used for this type of effluent are stabilization ponds and one or more of the activated sludge processes. As the major portion of the total wastewater organic loading is of high strength, because of the presence of the treating chemicals, it has been reported that the stabilization pond system is not very effective in reducing the BOD_5, COD and ammonia nitrogen concentrations(73,77). The available information indicates that sewage plant micro-organisms are incapable of degrading chemicals such as DEA and DEG. While MEA and ethylene glycol(EG) in high concentrations are toxic to the micro-organisms, these chemicals could be oxidized under operating conditions providing adequate dilution, pH control and aerobic activities.

It appears doubtful that effective biological degradation of the effluent by ordinary stabilization pond could be achieved in the northeastern British Columbia where the winter season lasts approximately six months. Because of low temperature and ice cover during the period of November to April, the biological activities in the pond at that period would be virtually nil. In addition, both the hydraulic and organic loadings vary from season to season due to natural precipitation variation, process upsets and annual equipment turnaround in summer.

There are several variations of the activated sludge system which have been used to treat this type of high organic wastewater. The basic process, shown schematically in Figure 26, involves transformation of the biodegradable waste into settleable solids by aeration in presence of micro-organisms. The organic matter in the waste is aerobically stabilized by the micro-organisms which draw required food and energy for growth from the oxidation of the waste. The settleable solids, mostly active masses of these micro-organisms, are settled in a sedimentation tank and disposed of as sludge, except for a small portion which is returned to the aeration tank as seed. The treated effluent is usually of low organic content.

Almost all types of activated sludge systems are affected adversely by both hydraulic and organic shock loadings. It is generally prevented by installing an equalization basin with adequate retention time upstream of the aeration tank to provide proper control of the wastewater feed. The anticipated removal efficiencies of pollutants like BOD_5, COD, sulphides, ammonia and suspended solids from oil refinery wastewaters are shown in Table 25. Because of the similarities in characteristics, the treatment of the gas processing

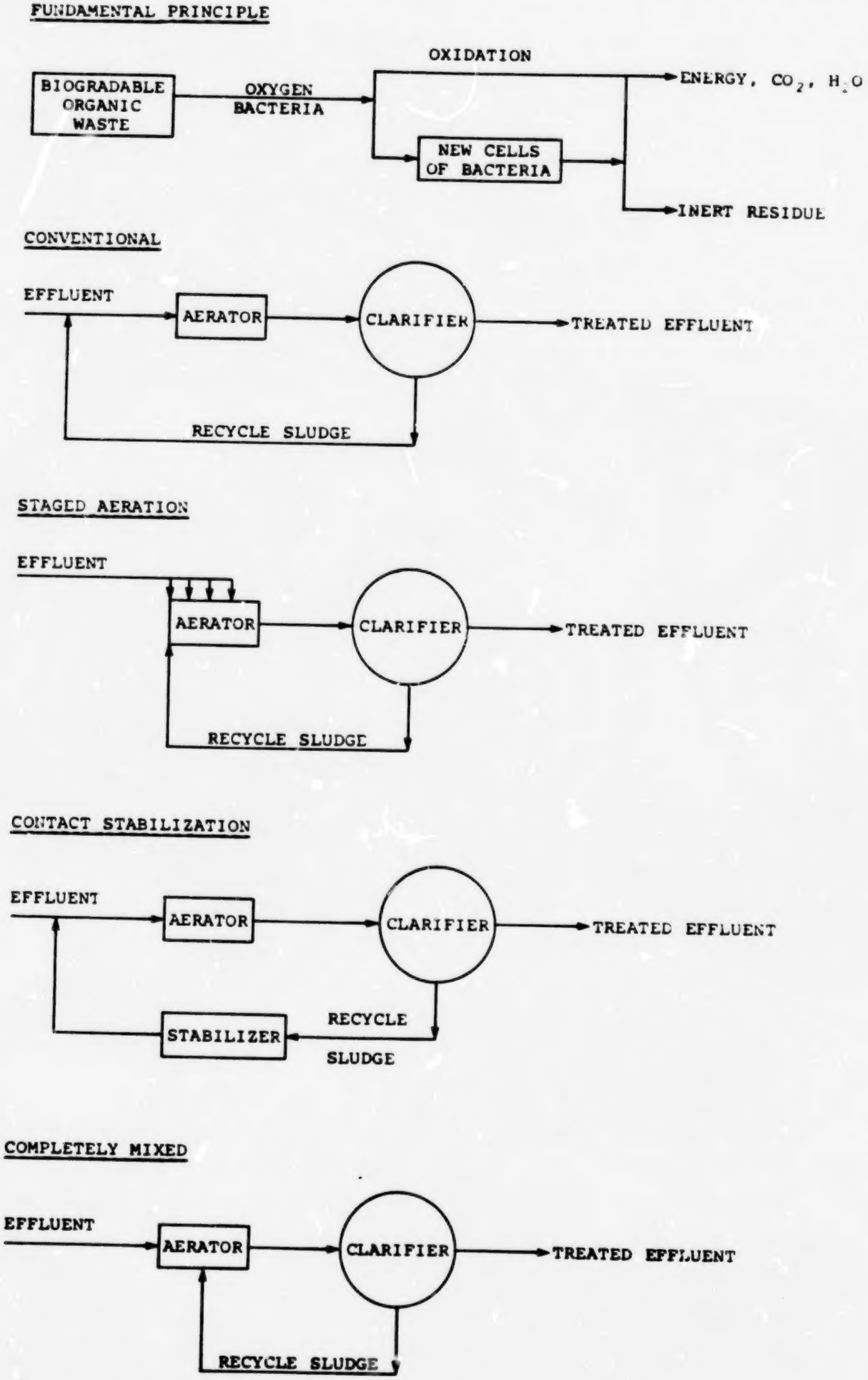

FUNDAMENTAL PRINCIPLE
BIOGRADABLE ORGANIC WASTE
OXYGEN BACTERIA
OXIDATION
ENERGY, CO_2, H_2O
NEW CELLS OF BACTERIA
INERT RESIDUE
CONVENTIONAL
EFFLUENT
AERATOR
CLARIFIER
TREATED EFFLUENT
RECYCLE SLUDGE
STAGED AERATION
EFFLUENT
AERATOR
CLARIFIER
TREATED EFFLUENT
RECYCLE SLUDGE
CONTACT STABILIZATION
EFFLUENT
AERATOR
CLARIFIER
TREATED EFFLUENT
STABILIZER
RECYCLE SLUDGE
COMPLETELY MIXED
EFFLUENT
AERATOR
CLARIFIER
TREATED EFFLUENT
RECYCLE SLUDGE

FIGURE 26: FLOW SCHEMATICS OF ACTIVATED SLUDGE TREATMENT MODES

TABLE 25

POLLUTANT REMOVAL EFFICIENCY OF WASTEWATER TREATMENT PROCESSES

Treatment	Pollutant Removal, % reduction							
	Oil & Grease	BOD_5	COD	$NH_3(N)$	Total $S^=$	Phenols	CN^-	Suspended Solids
Physical-Chemical								
Sour water stripper				69-96	96-99.8	10-30	75-90	
API separator	60-99	5-35	5-40					10-50
Air flotation without chemicals	50-90*	10-60	10-50	Reduced	Reduced	Reduced		50-90
Biological Method								
Activated sludge	50-80*	70-95	30-70	Reduced**	90-99	65-99**	65-99**	60-85
Aerated lagoons	70-90*	50-90	25-60		90-99			0-40
Trickling filters	50-80*	50-90	25-60		80-99			60-85
Oxidation ponds	50-90*	40-80	25-60		70-90			20-70
Tertiary Treatment* **								
Activated carbon		50-90	50-90		80-99	80-99		

* emulsified oil
** the range covering four processes
*** of secondary effluent

Source: (75, 76)

wastewaters by an activated sludge system is expected to produce similar degrees of reduction in the pollutants' level.

According to Milne(73), if the process wastewaters contain controlled amounts of MEA and EG, an extended aeration activated sludge system would be able to reduce the pollutants to acceptable levels. However, as DEA and DEG are not easily degraded biologically, it is doubtful whether the COD and $NH_3(N)$ levels of the effluent treated by this method alone would meet the standards. In Alberta at the Waterton plant of Shell Canada Ltd., where sulfinol is used as the desulphurization chemical, it was experienced that aerated lagoons and evaporation basins could not treat the high organic wastes adequately(77).

In this Province at the Fort Nelson plant, the original effluent treatment system with an activated sludge unit, as shown in Figure 27, was inadequate for the treatment of the combined process wastewaters and the sanitary effluent. In 1979 a modified treatment system was implemented incorporating an aeration lagoon downstream of the activated sludge unit (Figure 27). However, as evident from Figure 28, the treated effluent still contains consistently high level of $NH_3(N)$ and high levels of suspended solids and BOD_5 on occasions(78).

It appears from the foregoing that the presence of DEA in the process effluent may be a contributing factor to the high levels of $NH_3(N)$ and BOD_5 in the treated effluent. The practice of oil removal from the combined effluents of sanitary and process wastewaters may not be as efficient as to reduce the oil content to a satisfactory level whereby the biological treatment

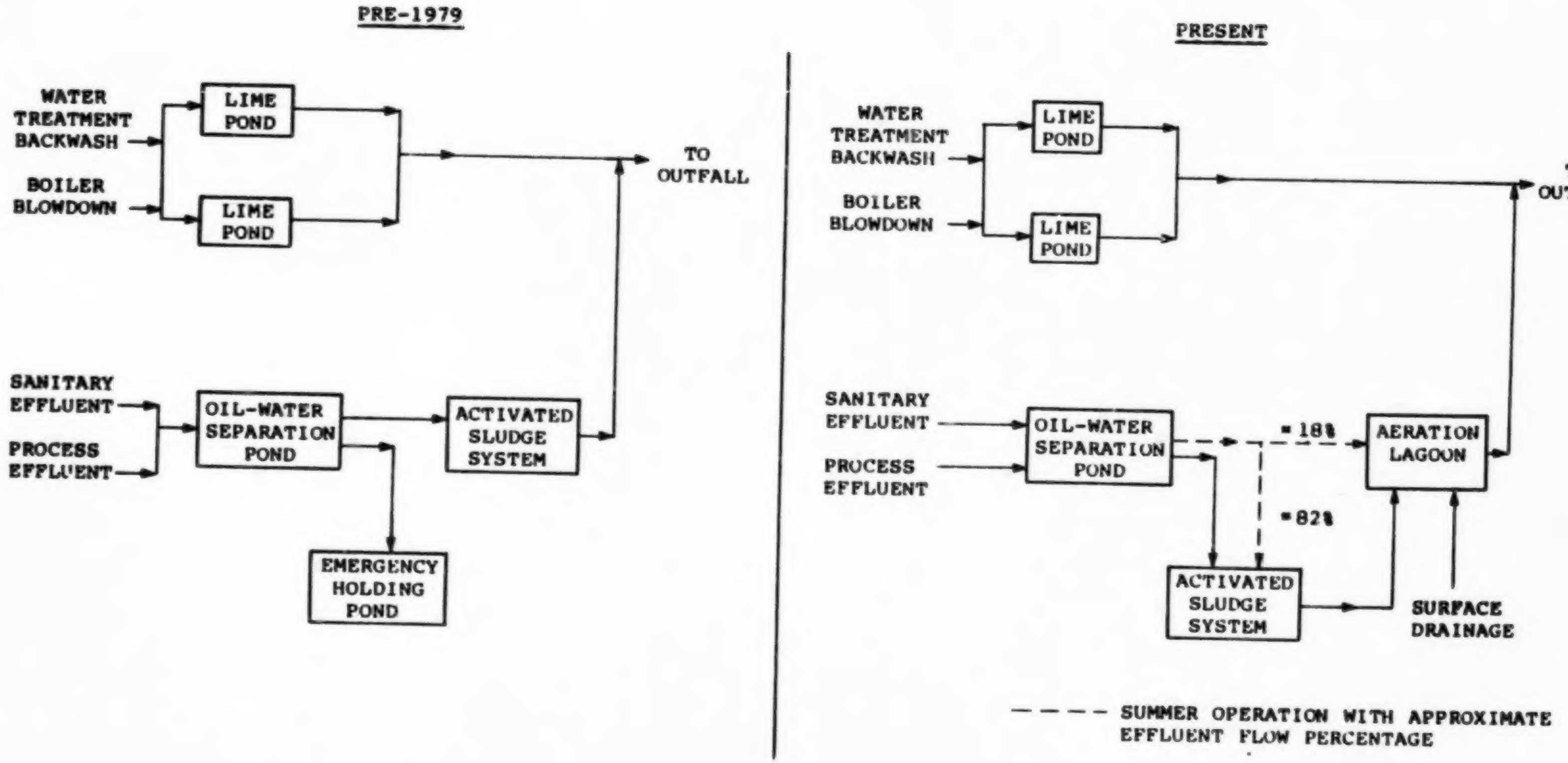

PRE-1979
WATER TREATMENT BACKWASH
BOILER BLOWDOWN
LIME POND
LIME POND
TO OUTFALL
SANITARY EFFLUENT
PROCESS EFFLUENT
OIL-WATER SEPARATION POND
ACTIVATED SLUDGE SYSTEM
EMERGENCY HOLDING POND
PRESENT
WATER TREATMENT BACKWASH
BOILER BLOWDOWN
LIME POND
LIME POND
TO OUTFALL
SANITARY EFFLUENT
PROCESS EFFLUENT
OIL-WATER SEPARATION POND
≈18%
≈82%
AERATION LAGOON
ACTIVATED SLUDGE SYSTEM
SURFACE DRAINAGE
SUMMER OPERATION WITH APPROXIMATE EFFLUENT FLOW PERCENTAGE

FIGURE 27: SCHEMATICS OF WASTEWATER TREATMENT SYSTEMS AT FORT NELSON PLANT

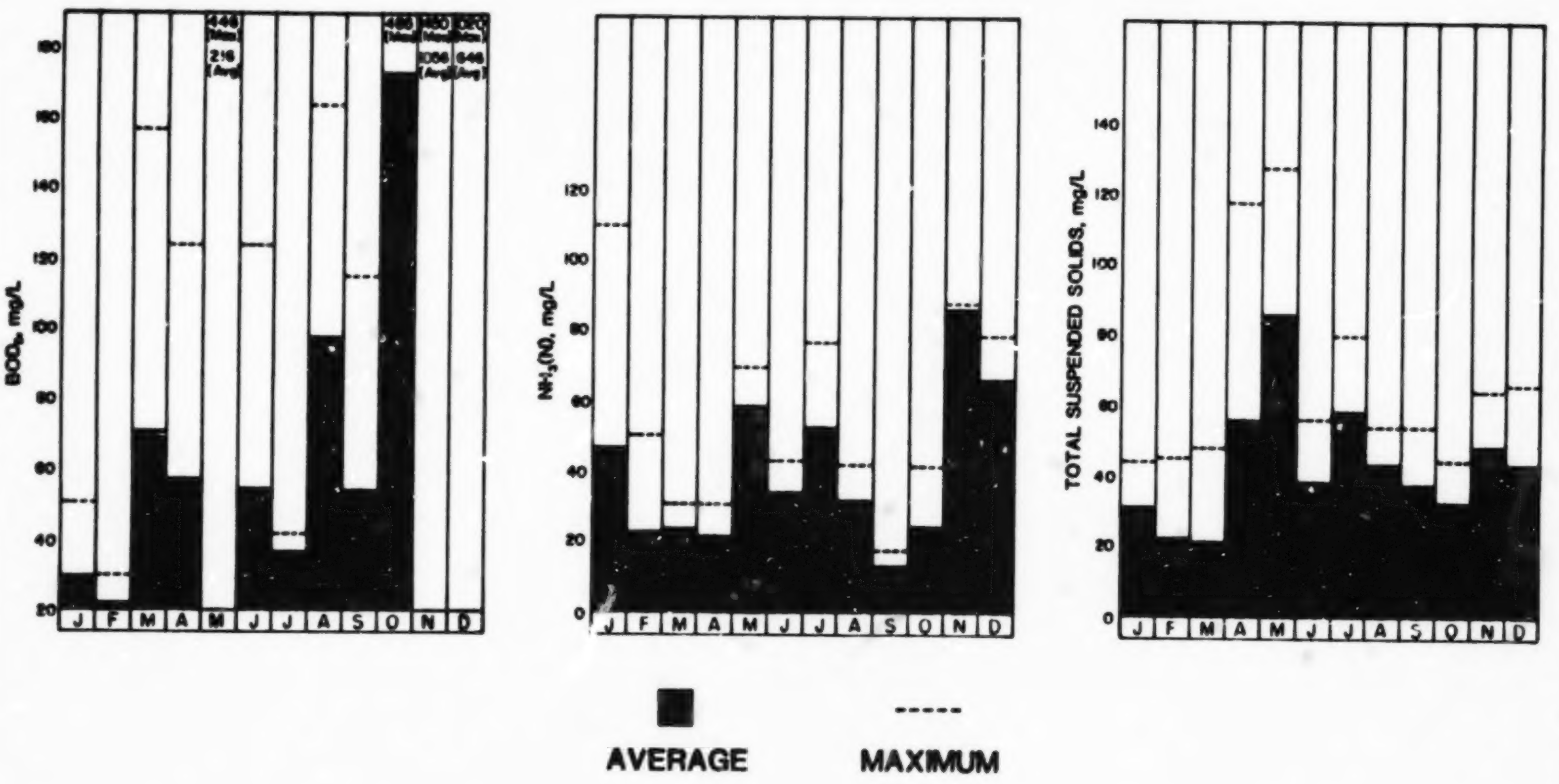


FIGURE 28: ANALYSIS OF FINAL EFFLUENT FROM FORT NELSON GAS PLANT - MONTHLY AVERAGE AND MAXIMUM VALUES FOR THE YEAR 1982

SOURCE: (78)

remains unaffected. As mentioned earlier, an excessive amount of oil tends to increase the suspended solids level in the biological treatment unit. Segregation of oily wastewaters and separate treatment system to remove both free and emulsified oil may have a positive effect on the final effluent quality. Additional chemical and physical methods can be used downstream of the biological system to produce an effluent of high quality.

6.2.4.2 Effluent Toxicity

In addition to the physical and chemical characteristics, the toxicity of the wastewaters is also considered to be of major importance. There are several components in the effluents, especially those originating from the processing areas, which are toxic individually or in combination with others to the aquatic biota. Depending on the particular treatment systems and chemicals used, even the treated effluent may show toxic behaviour in natural water-bodies. The components of known toxic nature and commonly present in the effluent are: sulphides, ammonia, suspended solids and oils. Other chemicals and metals are also likely to be present therein, which even in trace amounts may contribute towards the toxicity of the effluent because of the individual properties or by synergistic or antagonistic actions.

Because of the varied criteria which influence the toxic characteristics of the effluent, it would be appropriate to determine the toxicity of the treated effluent prior to its final discharge. Depending on the allowable toxicity level, it may be necessary to incorporate some kind of tertiary treatment facility to remove toxicity from the effluent.

As an example of an advanced wastewater treatment system, a brief description of the facilities used since June 1975 at the Waterton gas processing plant of Shell Canada Ltd. is provided here(77). The initial wastewater treatment facilities, consisting of sour water stripper, API separator, aerated lagoon and evaporation basin, were inadequate. The advanced system, shown schematically in Figure 29, is capable of producing final effluent quality shown in Table 26. A portion of the final effluent is recycled at the plant for preparing chemical solutions and for backwashing of filters. The remainder is discharged into a creek after ascertaining its toxicity by continuous monitoring.

6.2.5 Effluent Disposal

The manner in which various effluents can be disposed of would depend on factors such as plant location, availability of water and land, local climate, etc. The most common methods are:

- deep well injection,
- discharge of fully treated effluent into a waterbody,
- discharge of partially or fully treated effluent into exfiltration basins or on land by spray irrigation,
- incineration.

Depending on the circumstances any or a combination of the above methods may be used. However, certain degree of treatment of most effluents described above is necessary for efficient disposal without affecting the assimilative capacity of the receptor.

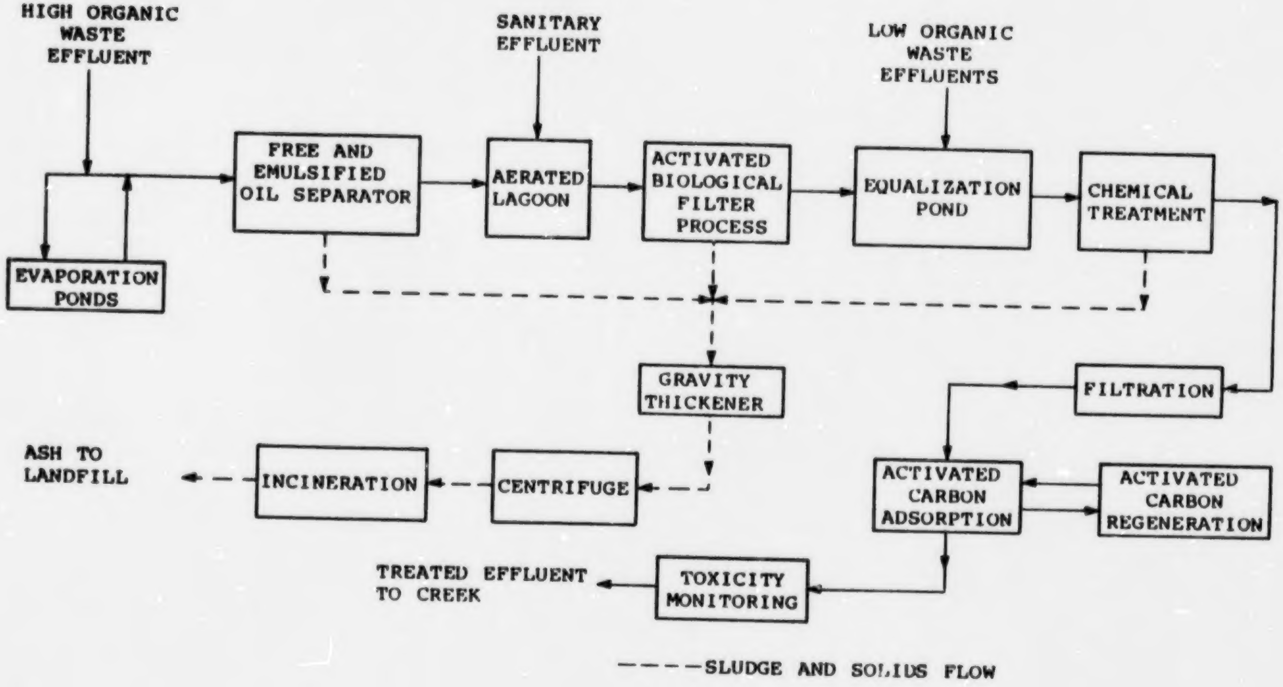

HIGH ORGANIC WASTE EFFLUENT
EVAPORATION PONDS
FREE AND EMULSIFIED OIL SEPARATOR
SANITARY EFFLUENT
AERATED LAGOON
ACTIVATED BIOLOGICAL FILTER PROCESS
LOW ORGANIC WASTE EFFLUENTS
EQUALIZATION POND
CHEMICAL TREATMENT
GRAVITY THICKENER
FILTRATION
ASH TO LANDFILL
INCINERATION
CENTRIFUGE
ACTIVATED CARBON ADSORPTION
ACTIVATED CARBON REGENERATION
TREATED EFFLUENT TO CREEK
TOXICITY MONITORING
----SLUDGE AND SOLIDS FLOW

FIGURE 29: **GENERAL FLOW SCHEMATICS OF WASTEWATER TREATMENT PLANT AT WATERTON GAS PLANT**

SOURCE: (77)

TABLE 26

EFFLUENT QUALITY GOAL OF SHELL CANADA LTD. AT WATERTON GAS PLANT, ALBERTA

Characteristics	Main Organic Waste	Final Effluent Goal
COD	3 200	25 - 50
Suspended Solids	160	0 - 10
Sulphides	50	0 - 0.2
Oil and grease	40	0 - 10
Ammonia (N)	5	0 - 5
Threshold odour number	>200*	<25*
Toxicity	Extremely toxic	Non-toxic to rainbow trout fingerlings

All concentrations are in mg/L unless otherwise stated.

*Units

Source: (77)

The deep well injection may be practised only after a complete evaluation of the geology, geochemistry, hydrology and seismic characteristics of the underground area, and ensuring that the effluent is not likely to affect the natural system. For disposal on land the degree of treatment required for the effluent would depend on the soil characteristics, hydrology and local climate in the area. The chemical compatibility of the soil and the effluent is essential for disposal by exfiltration and spray irrigation, as is the prevention of any ground or surface water contamination. The evaporation basins should preferably be lined with impermeable material and adequately sized to handle peak volumes of effluent. At the Pine River plant the original evaporation ponds are now considered inadequate and disposal of some of the effluents by spraying on the nearby land is practised.

Where incineration of wastewaters containing organic chemicals and oil is practised, it is essential that adequate flame temperature is maintained by burning supplementary fuel to prevent smoky combustion. Sufficient residence time for the combustibles in the combustion zone and mixing with the air are also prerequisites for complete oxidation of hydrocarbons and sulphides. The dispersion of resultant SO_2 and NO_x must be adequate to meet the ambient air quality objectives.

With a combination of judicious segregation of wastewater streams, proper design and operation of the treatment systems and good housekeeping practice, an effluent of quality can be produced which will cause a minimum damage to the receiving environment.

To minimize the effluent volume as well as for conservation of water, maximum reuse of the treated wastewater should be practised. The oil and condensates separated from the wastewater can be either hauled away for further processing or used as fuel at the plant.

6.3 Solid Wastes Management

The solid wastes generated from a natural gas processing plant are of the following types:

1. Oily sludge;
2. Inorganic sludges;
3. Organic solids and sludges; and
4. Spent catalysts and solid dehydrating agents.

Where activated carbon adsorption system is used for final treatment of the plant effluent, it is necessary to dispose of some inactive material from time to time.

Apart from the solids listed in item 4, the other three types of wastes can be disposed of either individually or together. The general practice has been, and still is at some installations, to discharge these wastes as landfills. However, because of the shortage of suitable disposal area, potential leaching effects on the receiving environment, and usable energy content of some of these wastes, alternative methods of disposal are also practised.

Oily sludges from free and emulsified oil removal systems contain, besides oil and water, rust, sand, mud and sometimes lime. These can be incinerated after thickening and removal of some water. The residue after incineration is smaller in volume and virtually inert in nature, and can be disposed of in a landfill. There are several designs of incinerator available for burning of this type of waste. The heat available from incineration can be incorporated into the plant energy system.

Inorganic sludges originate mainly from raw water treatment and boiler cleaning systems. The quality of these sludges depends on the plant supply water quality and the treating chemicals used (Table 21). These sludges are usually disposed of by landfilling. No environmental problem is likely to arise from such practice, provided no toxic chemicals and heavy metals are present.

Another sludge containing mostly diatomaceous earth is discharged during cleaning of the amine absorption unit. This is usually sent to a settling pond, and the solids are dredged as required and sent to a landfill or used in the plant yard.

The major volume of solid wastes in a gas processing plant originates from the biological treatment system. While a portion of the sludge is recycled to the activated sludge unit to maintain an adequate mixed liquor suspended solids, the remainder requires disposal. As the solids content of this type of sludge is usually very low (0.5 - 6.0%)(75), some conditioning of the sludge is usually necessary prior to disposal. Such conditioning essentially consists of thickening and dewatering of the sludge for ease of handling and disposal. In some cases, additional conditioning of the sludge by anaerobic digestion or wet oxidation or heat treatment may be required. The final disposal of the sludge is usually carried out by landfilling or incineration.

Alternative sludge conditioning and disposal methods are shown schematically in Figure 30. To aid in thickening of the sludge it is often necessary to use chemicals such as lime, ferrous sulphate, ferric chloride and polyelectrolytes. Dewatering of the sludge can be achieved by vacuum filtration, filter press, centrifugation or drying beds. Anaerobic digestion or wet oxidation of the thickened sludge helps the dewatering process. If the availability of land is not at a premium and it is suitable for accepting the conditioned sludge, then land disposal may be economical. However, depending on the site and the chemical nature of the sludge, certain preparation of the land by grading and lining may be necessary to prevent contamination of groundwater and surface water.

As an alternative, the sludge can be incinerated and the residue disposed of as a landfill. Incineration can be carried out in a fixed bed multiple hearth furnace or a fluidized bed system. The maximum temperature in the former type is about 980°C and that in the latter design is approximately 815°C(75). However, usually the incinerator is designed for each specific sludge, as its moisture content and calorific value determine its burning characteristics and the requirement for supplementary fuel. The flue gas from the incinerator is passed through a wet scrubber to control particulates and other gaseous pollutants.

In British Columbia, the solid wastes from the natural gas plants are used as landfills. At the Waterton plant in Alberta, all sludges - oily sludge, chemical treatment solids and biological solids - are mixed in a thickener(77). Thickened mixture is dewatered in a centrifuge, and then incinerated in a multiple hearth

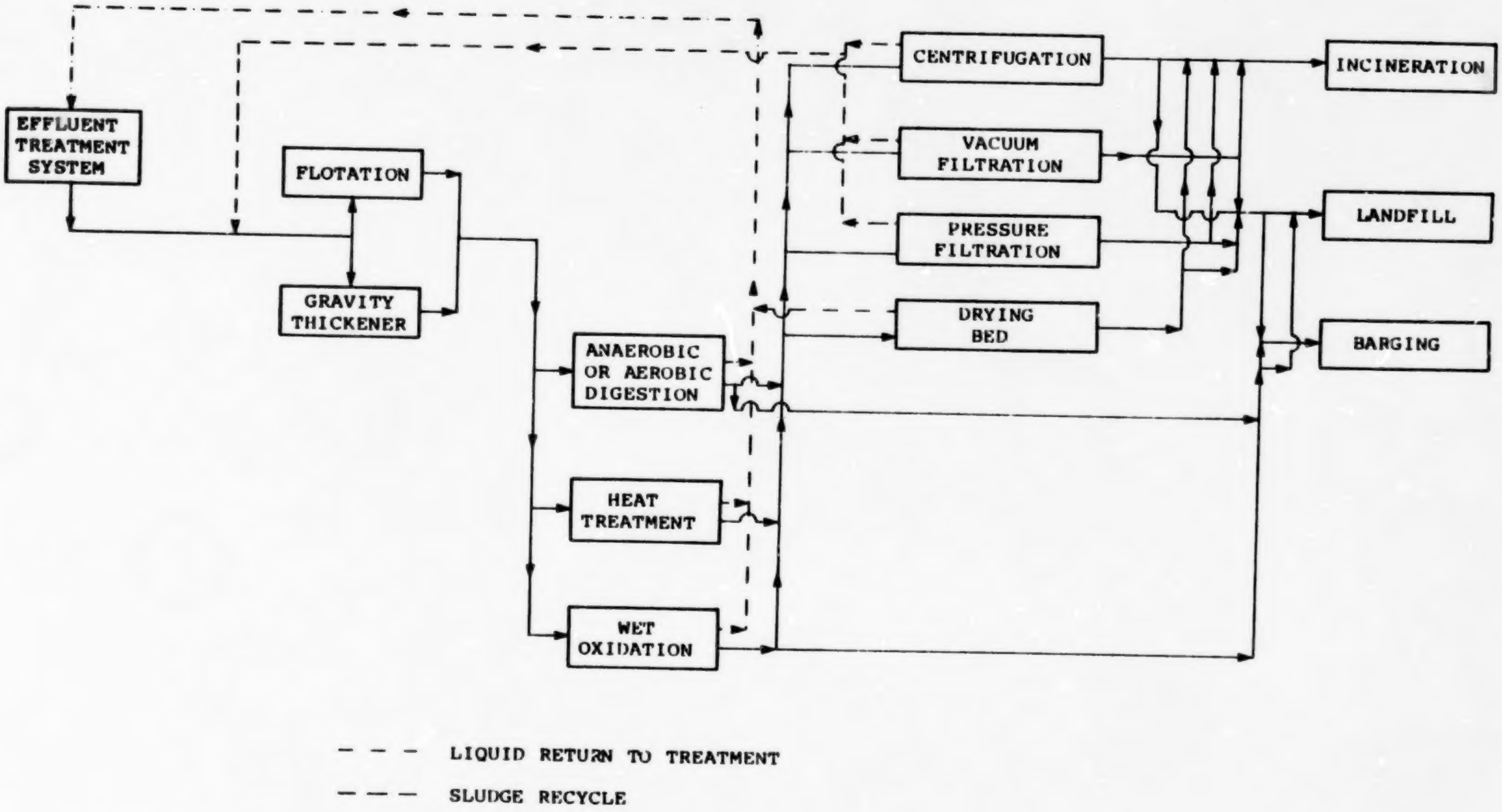

EFFLUENT TREATMENT SYSTEM
FLOTATION
GRAVITY THICKENER
ANAEROBIC OR AEROBIC DIGESTION
HEAT TREATMENT
WET OXIDATION
CENTRIFUGATION
VACUUM FILTRATION
PRESSURE FILTRATION
DRYING BED
INCINERATION
LANDFILL
BARGING
LIQUID RETURN TO TREATMENT
SLUDGE RECYCLE

FIGURE 30: ALTERNATIVES FOR SLUDGE CONDITIONING AND DISPOSAL

furnace at a temperature of about 925°C. Free oil collected from the oil separator is fired in the furnace as the supplementary fuel. The stack gas is passed through a wet scrubber to control particulates emission. The residue from the incinerator and the scrubber solids are disposed of in a landfill.

The spent catalysts, molecular sieves and any solid dehydrating agents require disposal or regeneration. The used molecular sieves and dehydrating agent are small in volume and easily regenerable, so their disposal in a landfill can be avoided. The spent catalysts from the sulphur plant are discharged during plant turn around at a rate of about 50 to 200 tonnes per 1 to 2-year cycle(22). The catalysts are mostly bauxite with some aluminum sulphate and iron. Although the spent material can be regenerated, the usual practice is to dispose of in a landfill.

Where the liquid effluent is treated in a bed of activated carbon for organics removal, the saturated carbon requires regeneration from time to time. This is usually achieved by heating the material in a furnace under reducing conditions at about 980°C. The desorbed gases from the activated carbon are burned at a higher temperature in an afterburner to ensure complete combustion. To minimize particulates emission the stack gas is scrubbed prior to disposal.

7. QUANTITIES OF WASTE DISCHARGES

The quantities of gaseous, liquid and solid waste discharges from a natural gas processing plant are basically site-specific. Various waste streams originating from a particular plant are both quantitatively and qualitatively dependent on the size of the plant, the type of unit operations employed, and the kind of control equipment used. There is very little information available in the literature on the specific quantities of individual waste streams generated per unit quantity of sour gas processed. After reviewing some relevant documents, including plant design and source testing data, an attempt has been made here to provide some general information on this subject matter.

7.1 Gaseous Emissions

In a typical gas-processing plant, as described under Section 6.1 and shown schematically in Figure 11, the emission sources can be grouped as those emitting products of combustion and others as sources of fugitive emissions. The first group of sources can be further categorized into:

(i) fuel burning equipment, such as heaters, boilers, compressor engines, turbines, etc.;
(ii) acid gas or sulphur plant tail gas burner; and
(iii) flare stack and pit.

To aid in estimating the rate of emission from fuel burning equipment, some relevant information on the combustion characeristics of natural gas and excess air requirements for various equipment is provided in Table 27. The calculated dry flue gas volumes per unit volume of natural gas and per unit of heat input at different excess air are illustrated in Figure 31. By knowing the fuel

TABLE 27

COMBUSTION CHARACTERISTICS OF NATURAL GAS

The following information is based on the typical anaylsis of sales gas given in Table 3. The standard conditions used for gas volumes are at 293.15K and 101.325kPa.

1. Calorific value: 36.96 MJ/m^3

2. 1 GJ = 27.06 m^3 of gas
   1 GJ = 19.46 kg of gas

3. Stoichiometric air requirement (assuming dry air) : 9.69 m^3/m^3 of gas
   262 m^3/GJ
   16.23 kg/kg of gas

4. Range of excess air requirement:

   Boilers - 5 - 20%
   Process heaters - 10 - 20%
   Reciprocating engines - 7 - 285%
   Gas turbines - 200 - 300%

Source: (3, 79-84)

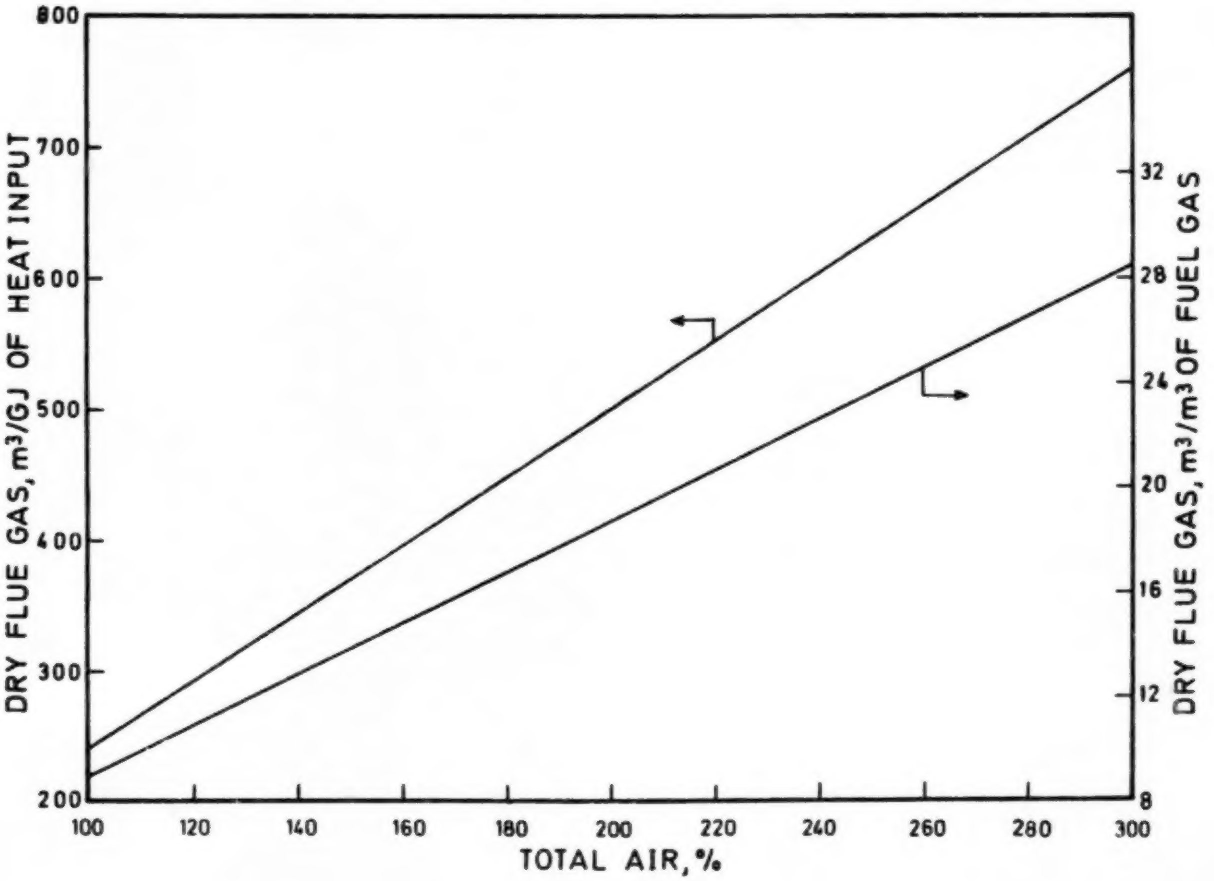


FIGURE 31: NATURAL GAS COMBUSTION CHARACTERISTICS

consumption rate and excess air used in an equipment, the stack gas volume can be estimated from Figure 31. Fuel consumption rates for packaged units, such as compressor engines, turbines and boilers, can be obtained from the name-plate information on the rated capacity. As far as process heaters are concerned, each unit is individually designed for the specific need.

Line heaters are sometimes necessary, particularly at wellhead, to prevent hydrate formation. In British Columbia, the heat input to such heaters appears to range from 1.65 to 21.12 GJ/h per 10^6 m^3 of well gas heated, depending on the gas characteristics and location(78).

Direct-fired heaters are used in some instances to regenerate glycol solutions used in dehydrators. According to Campbell(4), the rule of thumb about the glycol reboiler heat input is approximately 0.56 GJ/m^3 of the glycol pump capacity. The heat input to such heaters in existing facilities in this Province ranges from 0.6 to 2.4 GJ/h per 10^6 m^3/d of gas processed(78). For a dry desiccant regenerator the thermal efficiency may be assumed to be about 70% for the heater, and from its heat load the fuel consumption rate may be estimated(4).

In gas sweetening units in which the amine solution is regeneraed by direct-fired heaters, the heat input to the heater may be estimated from the following equation given by Campbell(4):

$$Q, \text{GJ/h} = 0.296 \times (\text{m}^3\text{/h of amine solution circulated})$$

This relation has been developed on the basis of the vapour entering the regenerator at 96°-107°C and leaving at 38°-54°C.

Packaged boilers are generally used to supply the steam necessary for various purposes in the plant. The boiler stack emission rate is related to the fuel consumption rate and the excess air usage. In a gas-fired boiler, the quantity of excess air used may vary from 5 to 20% of the stoichiometric air required for complete combustion. The fuel consumption rate in a boiler is dependent on its size and thermal efficiency. The latter ranges from 75% to 85% of the total heat input(79). The estimated natural gas consumption rate for various sizes of boilers at several efficiencies are shown in Figure 32.

In the gas-processsng industry reciprocating engines and gas turbines are used as compressors, pump drives and electricity generators, mostly as standby. The emission quantity from these combustors varies considerably because of several factors, but mainly due to their low efficiencies and variable but high rate of combustion air use. An accurate measurement of intake air and exhaust flue gas flow rates for these equipment is extremely complicated and difficult(82). The air intake by the engine or turbine is also dependent on the altitude and location of the equipment as well as on the method of air charging, e.g. naturally aspirated, blower scavenged or turbo-charged.

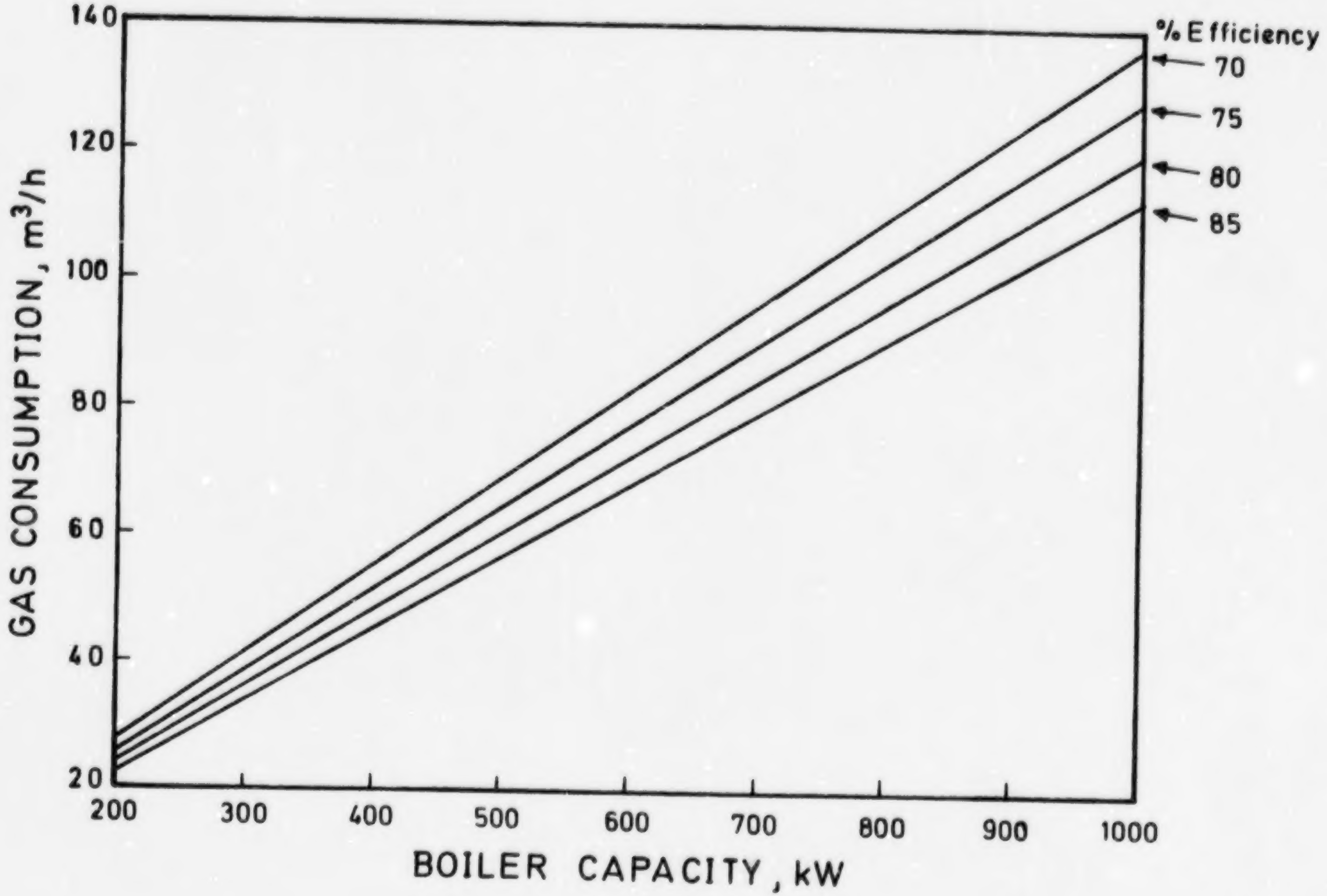


FIGURE 32: NATURAL GAS CONSUMPTION RATE FOR BOILERS OF VARIOUS CAPACITIES AT DIFFERENT EFFICIENCIES.

There is hardly any information on the exhaust gas flow rates from reciprocating engines and turbines. Some information is available on fuel consumption rate and air-to-fuel ratio for different types of equipment and this is summarized in Table 28. The fuel consumption rate is generally expressed as Brake-Specific Fuel Consumption (BSFC), which relates to actual energy output as opposed to the rated capacity. The latter is usually the capacity to generate maximum power under ideal conditions, and it also includes the amount expended in operating the engine accessories. From the information given in Table 28, at least some rough estimates of emission rates from these equipment could be obtained.

In the plants where sulphur recovery is practised, the major single source of emission is the sulphur plant tail gas incinerator or thermal oxidizer stack. Even in facilities without sulphur recovery but having large gas sweetening units, the disposal of the acid gas by incineration or flaring may be a significant source of emission. In both cases the final stack emission rate will primarily depend on the rate of acid gas generation and its quality. The latter, however, is site-specific and related to the rate of raw gas processed and its quality, the degree of gas sweetening, $H_2S:CO_2$ ratio in the acid gas, and the sulphur recovey process employed. The other factor which determines the stack emission rate is the method of disposal, namely flaring, thermal oxidation or incineration. Depending on the gas composition it may be necessary to use certain amount of fuel gas to ensure complete combustion, and this will affect the final emission rate. In an incinerator or thermal oxidizer an excess air up to 25% is also necessary, and if a waste heat boiler is used to recover some heat, an air blower is required to maintain the necessary pressure. Both these factors govern the final stack emission rate.

TABLE 28

FUEL CONSUMPTION RATE AND AIR/FUEL RATIO FOR RECIPROCATING ENGINES AND GAS TURBINES

Equipment	Rated Load kW	BSFC MJ/kw-h	Air/Fuel kg/kg
Reciprocating			
Engines			
Naturally-aspirated	298 - 1230	9.8-11.5	17.4-26.5
Blower scavenged	805	10.1	62.5
Turbocharged	530 - 5453	9.0-10.7	25.7-36.3
Gas Turbines	4623-19 962	15.6-17.5	N/A

N/A - not available
Fuel and air consumption rates are for emissions without NO_x control.

Source: (82, 84)

Considering the above aspects of acid gas disposal it becomes obvious that it is nearly impossible to establish a general formula for estimating the stack emission rate from this source per unit quantity of raw gas processed. A review of the plants in British Columbia shows that the rate of emission from the tail gas incinerator stack ranges from 0.40 to 0.75 m^3/m^3 of raw gas processed. In two plants where the acid gas is flared, the flaring rate varies from 0.017-0.05 m^3 of acid gas/m^3 of raw gas processed(78).

The flaring of gases and vapours in stacks and of condensates in pits is a common practice in the industry, but because of the method of burning these combustibles it is impossible to estimate the rate of emission of combustion products from these sources.

Fugitive emissions from gas processing plants are mostly volatile organic compounds (VOC) which are usually defined as non-methane and non-ethane hydrocarbons. In plants where LPGs are separated more VOC emission can be expected. In a recent study(85) by the U.S. Environmental Protection Agency it has been determined that fugitive emissions are related to the number of potential sources, but do not vary with capacity, throughput, age, pressure, or temperature. The fugitive emission rates for VOC and total hydrocarbons are shown in Table 29. The fugitive emissions of H_2S containing gases are usually well controlled because of its potentially toxic nature.

TABLE 29

BASELINE FUGITIVE EMISSION FACTORS FOR VOC AND TOTAL HYDROCARBONS FROM GAS PROCESSING PLANTS

(kg/d)

Emission Point	Emission Factor	95% Confidence Interval
Valves[a]	0.18 (0.48)	0.1 - 0.3 (0.2-1)
Relief valves[a]	0.33 (4.5)	0.007-8 (0.1-100)
Open-ended lines[a]	0.34 (0.53)	0.1-0.7 (0.2-1)
Compressor seals[a,c]	1.0 (4.9)	0.1-5 (0.7-30)
Pump seals[a]	1.2 (1.5)	0.5-3 (0.5-4)
Sampling Connections[b]		
Gas	0.016 (0.32)	
Liquid	0.085(0.085)	
Flanges and connections[a]	0.011 (0.026)	0.006-0.02(0.01-0.05)

Figures in parentheses are for total hydrocarbon emission values

[a] From tests carried out by Radian Corporation under contract to U.S.EPA.

[b] Liquid streams are assumed to be 100 percent VOC, sampled twice per month with a 1.96 litre purge. Gas streams are assumed to be sampled twice per shift with a 1 second purge through a 6.4 mm ID sample tube 15 cm long; 80% methane, 15% ethane, 5% VOC.

[c] Emission factors for compressors are based on EPA and API testing of emissions into the distance piece area from open frame compressors. The factors do not include emissions into the seal packing vent or into enclosed distance pieces. Therefore the emission factors given are probably understated substantially.

Source: (85)

7.2 Liquid Effluents

Estimates of some major liquid effluent flows generated in a hypothetical sour gas processing plant are shown in Table 30. However, as mentioned earlier, the actual wastewater generation rates are plant-specific. As shown in Table 31, the designed flow rates from two plants in British Columbia are vastly different. These differences are due to several factors some of which are listed below:

- the quality and volume of sour gas processed;
- the type of unit operations employed;
- the extent of air cooling and steam heating used; and
- housekeeping practices and recycle of usable materials.

The sanitary effluent flow rate can be reasonably estimated based on minimum daily quantities required per person. The recommended guidelines(87) for such flow estimation are as follows:

Office buildings - 0.0902 m^3/d/worker
Plants with showers - 0.0902 m^3/d/worker/shift
Plants without showers - 0.04546 m^3/d/worker/shift

The effluent generation rates from other sources are difficult to estimate unless the details of each unit process employed are known. As it is a common practice to combine several wastewater streams before final treatment, the actual rates of discharge of effluents to the receiving environment become harder to estimate. The

TABLE 30

ESTIMATED FLOW RATES OF SOME MAJOR WASTEWATER STREAMS GENERATED IN A HYPOTHETICAL SOUR GAS PROCESSING PLANT

($4.25 \times 10^6 m^3/d$ of raw gas containing 5% H_2S)

Wastewater Stream Source	Flow Rate m^3/d
Inlet separator water	24.6
Cooling tower and boiler blowdown	7.6
Plants and office floor washings	5.7
Washroom effluent	1.9
Filter washings	10.6
Dehydration process water	7.6
Reclaimer drain	0.2
Sulphur plant knockout water	3.0

Source: (86)

TABLE 31

DESIGNED RATES OF MAJOR EFFLUENT DISCHARGES FROM TWO GAS PROCESSING PLANTS IN BRITISH COLUMBIA

Effluent Stream	Effluent Volume (m^3/d)	
	Plant 1	Plant 2
Raw water treatment	70	22
Boiler blowdown	195	7[(a)]
Dehydration plant	19	(b)
Sweetening plant	91	(c)
Sanitary effluent including utility and shop drains	18	13
Sulphur storage area	0 - 454[(d)]	0 - 2940[(d)]
Raw gas process rate $10^6 m^3/d$	23	7
H_2S content, % (v/v)	1.5 - 4	11 - 16

(a) 90% of boiler blowdown is reused
(b) recycled
(c) incinerated
(d) surface runoff

Source: (78)

amount of effluents originating from a new plant can be reduced significantly by the use of available technology and water conservation by recycling and reuse.

7.2.1 Oily Wastes

Depending on the size of the operation there could be a significant quantity of usable oil generated from free oil separators. This oil is either used at the site or sold. It has been reported that at the Fort Nelson plant approximately 4.8 m^3 of such oil is recovered annually(88).

7.3 Solid Wastes

The quantities of solid wastes generated from a plant are site-specific, and the actual discharge quantities depend on the solid waste management practices.

At the Fort Nelson plant about 306 m^3/year of mainly raw water treatment sludge and 15 m^3/year of bio-sludge are discharged(88). No definite figures on spent catalysts and desiccants, scrap metal and other combustible garbage generated at the site are available.

At the Pine River plant on an average daily basis about 0.07 m^3 of wet sludge from the wastewater treatment plant is discharged. No spent catalysts or desiccants have yet been disposed of from the plant. It has been estimated that about 36 000 kg of desiccants every four years and 900 000 kg of spent catalysts every seven years would require disposal(88).

8. CONCLUSIONS

An attempt has been made to provide the necessary background information on the pollution control aspects of the natural gas processing industry. The available control technologies for various pollutants have been discussed with a major emphasis on the control of sulphurous emissions and sulphur recovery processes. Whenever appropriate the status of both process and control techniques in the existing plants in British Columbia has been discussed. The areas of potential improvement in the quality of certain discharges in view of the availability of alternative control technology have also been identified. It is emphasized, however, that additional investigations into specific problems at the existing installations are essential before implementing any changes to the process or control systems to achieve the desired objectives.

9. REFERENCES

1. "Natural Gas Processing Plants in Canada: Operations List 7", Mineral Resources Branch, Dept. of Energy, Mines and Resources, Ottawa, January, 1971.
2. "Commissioner Inquiry on British Columbia's Requirements, Supply and Surplus of Natural Gas and Natural Gas Liquids: Technical Report", George W. Govier, Commissioner, Victoria, British Columbia, May, 1982.
3. "A Combustion Handbook for Canadian Fuels, Vol. 2/Gaseous Fuels", Mines Branch Monograph 879, Energy, Mines and Resources Canada, 1973.
4. Campbell, John M.: "Gas Conditioning and Processing", Revised Edition, 1972, Published by John M. Campbell, Norman, Okla, U.S.A.
5. "Fuels Desulphurization": Informal Notes Prepared by Fuels Division, Air Pollution Control Directorate, Environment Protection Service, Ottawa, September 27, 1974.
6. Maddox, R.N.: "Gas and Liquid Sweetening", Second Edition, Campbell Petroleum Series, Norman, Okla., 1974.
7. Kohl, A.L. and F.C. Riesenfeld: "Gas Purification", Second Edition, Gulf Publishing Co., Houston, Tex., 1974.
8. Goar, B. Gene, The Oil and Gas Jl., July 12, 1971, 75, and July 19, 1971, 84.
9. Blake, R.J., The Oil and Gas Jl., January 9, 1967, 150.
10. Anon., Oilweek, June 23, 1975, 53.

11. Maddox, R.N. and Michael D. Burns, The Oil and Gas Jl., September 9, 1968, 130.
12. Goar, B. Gene, The Oil and Gas Jl., June 30, 1969, 117.
13. Klein, J.P., Oil and Gas International, September 1970, 109.
14. Grekel, H., The Oil and Gas Jl., July 20, 1959, 76.
15. Haines, H.W., G.A. VanWielingen, and Palmer, G.H., Petrol. Refiner., 40, 4, April 1961, 123.
16. Beach, F.K., Can. Oil and Gas Industries, July 1962, 41.
17. Riesenfeld, F.C. and C.L. Blohm, Hydrocarbon Processing and Petroleum Refiner, 41, 4, April 1962, 123.
18. Maddox, R.N. and Michael D. Burns, The Oil and Gas Jl., January 22, 1968, 91.
19. Townsend, F.M. and L.S. Reid, Petrol. Refiner, 37, 11, November 1958, 263.
20. Ellwood, P., Chem, Engrg., 71, 15, July 20, 1964, 128.
21. Colley, D.G., E.M. Berlie, and M.J. Zelensky, Paper presented at PNWIS-APCA Annual General Meeting, Vancouver, Canada, November 1982.
22. Klemm, R.F., "Environmental Effects of the Operation of Sulfur Extraction Gas Plants", Report submitted to Environment Conservation Authority, Alberta, September 1972.
23. "Pollution Abatement Review With Respect to Heavy Crude Oil Upgrading, Synthetic Crude Use at Refineries, Increased Natural Gas Processing and Liquefied Natural Gas Facilities", Report prepared for Department of Supply and Services, and Environment Canada by RTM Engineering Ltd., Calgary, Alberta, August 1982.

24. Proceedings of Alberta Sulphur Gas Research Workshop III, held at The University of Alberta, Edmonton, Alberta, November 17-18, 1977.
25. Goar, B.G. and T. Arrington, The Oil and Gas Jl., June 26, 1978, 160.
26. Berlie, Elmer M., Paper presented at the Fourth Joint Chemical Engineering Conference of A.I.Ch.E./C.S.Ch.E., Vancouver, B.C., September 1973.
27. Grekel, H., L.V. Kunkel, and R. McGalliard, Chem. Engrg. Prog., 61, 9, 1965, 70.
28. Beers, W.S., U.S. N.T.I.S. Rep. No. PB 220 376, 1983.
29. Moyes, A.J. and S. Vasan, The Oil and Gas Jl., September 2, 1974, 56.
30. Vasan, S., The Oil and Gas Jl., January 2, 1978, 78.
31. Miller, S.G. and R.D. Robuck, Jl. Petro. Techno., 24, 1972, 545.
32. Anon., Chem, Engrg., 79, 19, September 19, 1960, 166.
33. Jennett, E., The Oil and Gas Jl., April 30, 1962, 72.
34. Anon., The Oil and Gas Jl., August 7, 1978, 92.
35. Gamson, B.W. and R.H. Elkins, Chem. Engrg. Prog., 49, 4, 1953, 203.
36. Barry, C.B., Hydrocarbon Processing, 51, 4, 1972, 102.
37. Anon., The Oil and Gas Jl., July 24, 1978, 46.
38. "Control Techniques for Sulfur Oxide Air Pollutants", U.S. Dept. of H.E.W., N.A.P.C.A., Washington, D.C., January 1969, 79.

39. Saltzman, R.S. and E.B. Hunt, Paper presented at I.S.A. Meeting, San Francisco, Ca., 1972.

40. Bonifay, P. and R. Dutriau, Gas Processing/Canada, January-February, 1972, 38.

41. "Compilation of Air Pollutant Emission Factors Third Edition (Including Supplements 1-7) Supplement No. 10, U.S. E.P.A., PB 80-199045, February 1980.

42. Anon., The Oil and Gas Jl., August 28, 1978, 160.

43. Goddin, C.S., E.B. Hunt and J.W. Palm, Hydrocarbon Processing, 53, 10, 1974, 122.

44. Krill, Helmut and Klaus Storp, Chem. Engrg., July 23, 1973, 84.

45. Rowland, Les, Oilweek, June 28, 1971, 18.

46. Cameron, L.C. and P. Grancher, "Sulfreen Process: Status of Development", Paper presented at 53rd Annual Convention of GPA, Denver, Colo., March 25-27, 1974.

47. Martin, J.E. and Guynot, G., Gas Processing/Canada, November-December 1971, 26.

48. Cameron, L.C., The Oil and Gas Jl., June 24, 1974, 110.

49. Andrews, John W., et al, JAPCA, 26, 7, July 1976, 664.

50. Barthel, Y., et al, "The Technology of Claus Unit Tail Gas Cleanup with the IFP process", Paper presented at 53rd Annual Convention of GPA, Denver, Colo., March 25-27, 1974.

51. Conser, R.E. and R.F. Anderson, The Oil and Gas Jl., October 29, 1973, 81.

52. Anon., The Oil and Gas Jl., September 11, 1978, 88.

53. Bailleul, M.R., et al, "Pollution and Sulfur Plants", Paper presented at 49th Annual Convention of GPA, Denver, Colo., March 17-19, 1970.

54. Anon., Chem. Engrg., September 30, 1974, 56.

55. Naber, J.E., J.A. Wesselingh, and W. Groenendaal, Chem. Engrg. Prog., 69, 12, 1973, 29.

56. Naber, J.E., et al, Energy Processing/Canada, September-October 1973, 32.

57. Swaim, C.D., "The Shell Claus Off-gas Treating (SCOT) Process", Paper presented at the 167th Amer. Chem. Soc. Meeting, Los Angeles, Ca., March 31-April 5, 1974.

58. Beavon, D.K., R.H. Hass, and B. Muke, The Oil and Gas Jl., March 12, 1979, 76.

59. Beavon, D.K. and R.N. Fleck, "Beavon Sulpur Removal Process for Claus Plant Tail Gas", Paper presented at the 167th Amer. Chem. Soc. Meeting, Los Angeles, Ca., March 31-April 5, 1974.

60. Beavon, D.K., Chem. Engrg., December 13, 1971, 71.

61. Crow, J.H. and J.C. Bauman, Hydrocarbon Processing, 53, 10, 1974, 131.

62. Laengrich, A.R. and W.L. Cameron, The Oil and Gas Jl., March 27, 1978.

63. Potter, B.H. and T.L. Craig, Chem. Eng. Prog., 68, 8, 1972, 53.

64. Informational Letter, IL 80-24 issued by Energy Resources Conservation Board and Alberta Environment, The Province of Alberta, November 4, 1980.

65. Information submitted by Westcoast Transmission Co. Ltd. with the application for Pollution Control Permit, June 13, 1979.

66. McConnell, Anne E., "A Monitoring Program and Some Design Proposals for Pine River Gas Plant Spray Irrigation Project", Westcoast Transmission Co. Ltd., September 10, 1982.

67. "Soil Biochemistry" by A. McLaren and J. Skujins, 1971, Marcel Dekker, N.Y.

68. "Introduction to Soil Microbiology", by M. Alexander, 1967, John Wiley and Sons, Inc., N.Y.

69. Kemp, E., J.B. Hyne, and W. Rennie, Int. Jl. Sulphur Chemistry, 1, 1, 1971, 69.

70. Rennie, W.J., D. Dunay, and J.B. Hyne, Alberta Sulphur Research Quarterly Bulletin, Vol. VIII, Nos. 1 and 2, 1971.

71. Duffy, J.J. and A. Smandych, Proc. of Alberta Sulphur Gas Research Workshop III, Held at The Univ. of Alberta, Edmonton, November 17-18, 1977.

72. "Development Document for Effluent Limitations Guidelines and Standards for the Steam Electric Point Source Category (Proposed)", EPA-440/1-80/029-b, September 1980, U.S.E.P.A., Washington, D.C.

73. Milne, Kevin C., Energy Processing/Canada, November-December 1974, 24-47, and January-February 1975, 30-34.

74. References quoted in "Coal Conversion Control Technology Volume 1. Environmental Regulations; Liquid Effluents", EPA-600/7-79-228 a, October 1979, U.S.E.P.A., Washington, D.C.

75. Eckenfelder, Jr., W. Wesley, "Water Quality Engineering for Practicing Engineers", Barnes and Noble, Inc., N.Y., 1970.

76. Grant, W.J., Paper presented at Petroleum Refinery Technology and Pollution Control Seminar, Sponsored by E.P.S. and P.A.C.E., Edmonton, Alberta, October 2-4, 1974.

77. Christensen, Douglas R., and Conn, Barry R., Paper presented at the Canadian Natural Gas Processing Association Meeting, Calgary, Alberta, June 1976.
78. B.C. Ministry of Environment, Waste Management Branch, Permit File Nos. PA-1555, -1780, -2803, -3457, -5056, -5057, -5058, -5151, -5364 & -5365; PE-1472 & -5615; PR-2376 & PR-6321.
79. Fryling, Glenn R., "Combustion Engineering", Revised Edition, Combustion Engineering Inc., N.Y. 1967.
80. Blokker, P.C., "Survey on nitrogen oxide emissions from stationary sources and techniques for reducing these emissions", Report No. 4/75, March 1975, Stichting CONCAWE, The Hague, Netherlands.
81. Information from Cleaver-Brooks' Co. on CBH Packaged Boilers.
82. "Air Pollution Emissions and Control Technology: Thermal Power Generation Industry. Vol. 1, Internal Combustion Engines", Economic and Technical Review Report EPS 3-AP-76-5, APCD, EPS, Fisheries and Environment Canada, November 1977.
83. "Control Techniques for Nitrogen Oxide Emissions from Stationary Sources", U.S. Dept. of H.E.W., N.A.P.C.A., Washington, D.C., March 1970, 5-6.
84. "Stationary Internal Combustion Engines. Standards Support and Environmental Impact Statement, Vol. 1: Proposed Standards of Performance", Draft EIS, EPA-450/2-78-125a, July 1979, U.S. E.P.A., Washington, D.C.
85. "Equipment Leaks of VOC in Natural Gas Production Industry - Background Information for Proposed Standards", EPA-450/3-82-024a, December 1983, U.S.E.P.A., Washington, D.C.

86. Younger, A.H., Lecture notes for the course on Natural Gas Processing, Chemical Engineering Dept., University of Calgary, July 1984.
87. Information supplied by the Municipal Section, Technical Services, Waste Management Branch, Ministry of Environment, July 1984.
88. Information obtained from Westcoast Transmission Company Ltd., June 1984.